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APPLICABILITY OF REDUCTION TO SULFUR TECHNIQUES TO THE DEVELOPMENT OF NEW PROCESSES FOR REMOVING SO_2 FROM FLUE GASES

Final Report - Volume I

Phase I of Contract PH-22-68-24

For:

New Process Development Section
Division of Process Control Engineering
National Air Pollution Control Administration
U. S. Department of Health, Education, and Welfare

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This is Volume I of a two-volume report which summarizes the results and conclusions of a study of SO2 reduction processes. This volume I covers phase I activities, with the aim of establishing the state-of-the-art of reduction to sulfur techniques. Based on a comprehensive literature survey, thirty case studies were worked up covering several types of sulfur oxide stack emissions and several reductants. Each case was based on the best information available, and use of updated technology in devising the process sequence. Flow sheets, operating parameters and economics are reported in this volume.

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FOREWORD

This two-volume report summarizes the results and conclusions made by the Industrial Chemicals Division of Allied Chemical Corporation for the National Air Pollution Control Administration under Contract No. PH-22-68-24.

Volume I covers Phase I activities performed between June 1, 1968 and July 31, 1969. The objective of Phase I was to establish the state-of-the-art of reduction to sulfur techniques. Based on a comprehensive literature survey, thirty case studies were worked up covering several types of sulfur oxide stack emissions and several reductants. Each case was based on the best information available, and use of updated technology in devising the process sequence. Flow sheets, operating parameters, and economics are reported in this volume.

Volume II covers Phase II activities performed between August 1, 1969 and September 30, 1970. Four of its five sections deal with experimental studies designed to optimize process conditions and to confirm the validity of assumptions made in the Phase I studies. These sections cover (1) Claus Process Kinetics, (2) Intermediate Reactor Studies, (3) Low Temperature Claus Process Studies, and (4) Strong SO_2 Reduction Studies. The fifth section is a Phase I type study on the use of dimethylaniline to gather the SO_2 from smelter gases and deliver a concentrated SO_2 gas to a reduction process.

The contract work reported herein covers an important approach to SO_2 pollution control. It shows the reduction to sulfur technique to be a viable vehicle. By establishing certain process applications to be either technically or economically untenable, it narrows the area to a few preferred routes. An example is the handling of smelter gas by a gathering process, with subsequent reduction of the concentrated SO_2 to sulfur. Laboratory investigations have optimized operating conditions for selected process steps, specifically the Intermediate Reactor and the Claus units, that are common to almost all reduction flow sheets.

Although many departments within the Allied Chemical complex contributed to this work, Industrial Chemicals Division's R & D staff held prime responsibility for its planning, direction, and completion. The following individuals had major time participation in this effort: Mr. C. A. Bernales, Intermediate Reactor and Strong Gas Studies in Phase II; Mr. S. B. Boucher, engineering support in Phase I; Mr. R. L. Burrell, Low Temperature Claus Studies in Phase II;

FOREWORD
(Continued)

Mr. R. H. Edgecomb, process engineering in Phase I; Mr. G. B. Falk, economic evaluations in Phases I and II; Mr. T. S. Harrer, engineering support in Phase I; Mr. R. S. Park, process engineering in Phases I and II; Dr. L. P. Sharma, general consultant and Strong Gas Studies in Phase II; Mr. R. L. Sturtevant, general consultant in Phase II; Dr. S. N. Subbanna, Claus Kinetic Studies in Phase II; and Mr. A. W. Yodis, Project Director of Phases I and II.

The cooperation of our NAPCA Project Officer, Mr. G. L. Huffman, is especially appreciated.

CONTENTS

Page

1.0 Introduction -- 1

2.0 Approach -- 2

  2.6 Standard Flue Gases ---------------------------- 3

3.0 Summary of Results -- 4

4.0 Limitations on Estimates -- 4

  4.2 Fixed Capital ---------------------------- 4

  4.3 Working Capital ---------------------------- 4

  4.4 Feed Gas Value ---------------------------- 4

  4.5 Methane and Coke Values ---------------------------- 5

  4.6 Return on Investment ---------------------------- 5

  4.7 Steam Credits ---------------------------- 5

  4.8 Electricity ---------------------------- 5

  4.9 Combusted Flue Gas ---------------------------- 5

  4.10 Other Cost Bases ---------------------------- 6

5.0 Economic Commentary -- 6

  Summary of Results, Table III ---------------------------- 5A

  Summary of Results, Table IV ---------------------------- 5B

  5.1 Coke ---------------------------- 6

  5.2 Natural Gas ---------------------------- 6

  5.3 Pure H_2 and Pure CO ---------------------------- 7

  5.4 Reformed Methane ---------------------------- 7

  5.5 Hydrogen Sulfide ---------------------------- 7

  5.6 Low Temperature Claus ---------------------------- 7

  5.7 Cost Effects SO_2-O_2 Concentration ---------------------------- 7

  5.8 Power Stack Gas ---------------------------- 8

  5.11 Smelter Gases ---------------------------- 9

  5.12 Upstream Gathering ---------------------------- 9

  5.13 Future Work ---------------------------- 10

Page

6.0 Process Studies ---------- 10

6.1 Technology related to the production of feed gases ---------- 10

6.2 Cooling undiluted converter gas 2A ---------- 12

7.0 Hot Coke Reduction ---------- 13

7.1 Process Background ---------- 13

7.2 Discussion of Thermochemistry, Equilibria and Kinetics ---------- 13

7.3 Coke Reduction Process Studies ---------- 16

7.4 Factors for Scale-up and Equipment Sizing ---------- 17

7.5 The Regenerator System ---------- 18

7.6 The Electrostatic Precipitator ---------- 18

7.7 The Intermediate Catalyst Converter ---------- 18

7.8 The Claus Unit ---------- 19

7.9 Commercial Plant Projection ---------- 19

8.0 Reduction of SO_2 with Methane ---------- 19

8.1 Background ---------- 19

8.2 Primary Reactions ---------- 19

8.3 Process Profiles ---------- 24

8.4 Process Chemistry ---------- 25

8.5 Yushkevich - Reduction of Gas No. (90% SO_2) ---------- 31

8.6 West (TGS) Process - Gas No. 1 ---------- 34

8.7 Kulcsar Process - Gas No. 1 ---------- 37

8.8 Asarco Process - Gas No. 1 ---------- 40

8.9 Reduction of Gas No. 2A(Smelter gas) ---------- 42

8.10 West Process ---------- 43

8.11 Reduction of Gas No. 2B (Smelter gas) ---------- 43

8.12 West Process ---------- 46

8.13 Kulcsar Process ---------- 46

Page

Incomplete information:

8.14 Reaction Kinetics ---------------------------------- 46
8.15 Materials of Construction -------------------------- 47
8.16 Intermediate Converter------------------------------ 47
8.17 Catalyst Life -------------------------------------- 47
9.0 Reformer Gas Reduction ------------------------------ 48
9.1 Introduction -- 48
9.6 Reduction of Gas No. 2A (Smelter gas) --------------- 48
9.7 Conclusions --- 50
10.0 Reduction with H_2S ------------------------------ 51
10.1 Introduction --------------------------------------- 51
10.7 H_2S Generation ---------------------------------- 51
10.8 H_2S Generation from CH_4 and SO_2 ----------- 52
10.9 H_2 Generation from CH_4 and S_2 ------------- 53
10.12 Reduction of SO_2 with H_2S - Normal Claus ---- 54
11.0 Reduction of SO_2 with H_2S - Low Temperature Claus ------ 56
11.1 Introduction --------------------------------------- 56
11.4 Studies on Catalysis and Absorption ---------------- 57
11.5 Low Temperature Claus Process Studies -------------- 64
11.6 Case 1 - Type 2B Smelter Gas ----------------------- 65
11.7 Case 1a - Type 2B Smelter Gas ---------------------- 67
11.8 Case 1b - Type 2B Smelter Gas ---------------------- 68
11.9 Case 1c - Type 2B Smelter Gas ---------------------- 70
11.10 Case 1d - Type 2B Smelter Gas --------------------- 70
12.0 Low Temperature Claus Applied to Power Stacks ------ 72
12.3 Design Problems ------------------------------------ 72

Page

Low Temperature Claus Applied to Power Stacks (Continued)

12.4 Case 2 - Chain Grate Moving Bed Converter ---------- 74
12.5 Case 2B - Fixed Bed (Switch Reactors) -------------- 81
12.6 Case 2A - Moving Bed Converter Applied to Stack Gas ------------------------------------ 86
12.7 Comments on Sulfur Loading -------------------------- 86
12.8 Case 6 - Low Temperature Claus Applied to Type 2C Smelter Gas ------------------------------ 87

13.0 Reduction of SO_2 with Metallic Sulfides ------------------ 90
14.0 Equilibrium Calculations - Thermodynamic Background ------ 90
Low Temperature Claus Equilibrium Studies
15.0 Appendix -- 104
16.0 Bibliography and References ------------------------------ 283

Note: Exhibit material appearing in earlier monthly reports is given with a prefix number designating the number of the report in which it first appeared. The same applies to tables in the text. Tables appearing for the first time in this report are given in Roman numerals. Exhibits appearing for the first time in this report are prefixed with number 11-.

Final Report - Volume I

PH 22-68-24

1.0 Introduction

1.1 Practically all the world's supply of copper, nickel, zinc and lead is obtained from sulfide ores. When these ores are roasted or smelted, sulfur dioxide is evolved. So, too, when sulfur-bearing coals and oils are burned for heat or power, SO_2 is the end-product in the gas discharged to the atmosphere. Reduction of SO_2 to elemental sulfur is one useful approach to the abatement of the resulting pollution. Under subject contract, Allied Chemical Corp. has, since June 1968, been studying various SO_2 reduction processes.

1.2 At the outset, the task was divided into three phases:

Phase I - A comprehensive literature survey to select and evaluate potentially useful techniques.

Phase II - Laboratory or bench-scale experimentation to fill gaps in the data gleaned from the literature, to test the assumptions on which Phase I technical and economic evaluations rest, and to develop or optimize selected reduction processes.

Phase III- An engineering and economic study of processes surviving Phases I and II which is sufficiently detailed to determine whether further development and ultimate commercialization is justified.

1.3 This is an Interim Report covering the accomplishments of Phase I studies over the 14 month period ending July 31, 1969. The possibilities of this mode of control are by no means exhausted as of this writing. It is, therefore, intended to continue Phase I type evaluations as new ideas arise, even as Phase II studies move forward.

1.4 The potential sulfur recoverable from the enumerable sources is vast. As more and more of these now wasted resources are tapped, elemental sulfur will be seen as the preferred product over sulfuric acid by reason of its easy storability and economy in shipping - factors essential to the preservation of an orderly market. While conversion to acid may in certain circumstances provide a viable solution, reduction to sulfur provides a more general, long-term solution, which at once conserves a resource and avoids converting an undesirable exit gas into an effluent or solids waste disposal problem.

2.0 Approach

2.1 A broad search of the literature was the first step. Given in the Appendix is a list of 625 references found in this search and grouped according to reductant concerned. Those references which proved most useful on critical examination have been asterisked. Only a handful of practical reductants and reduction processes were found. Very few had been advanced beyond laboratory scale investigation to pilot plant or commercial practice.

2.2 The plan, as it evolved, discarded the concept of finding ready-made, well-defined processes in the literature. Instead, the plan relied on devising a practical composite process for each situation, using updated technology and the best data available, together with judgments based on direct experience.

2.3 To implement this, the literature was classified according to the reductant used. Only four major reductant classes were considered:

- solid carbon, specifically coke
- gaseous hydrocarbons such as methane
- hydrogen sulfide
- reformer or producer gas (mixtures of CO and H_2)

2.4 Several reductants were rejected as contenders, among them pure hydrogen, pure CO, coal, petroleum coke and certain metal sulfides, among them FeS, PbS and ZnS. Of these sulfides, FeS was the most favorable, but still not economically viable (See Section 12.0). Hydrogen appeared to have no special virtue over CH_4 or reformer gas, and is considerably more costly. While experimental work might have clarified the potential of some of these reductants, or allowed substantial departures from the processes described, no experimental work was authorized for any part of our Phase I effort. Coal, for example, may be feasible, but not in the kind of process and equipment reported by Cominco (76) In effect, one would be operating a coking process, contaminating the product sulfur with coal tar, ash and unburned carbon.

2.5 Petroleum coke is of two kinds. Delayed green coke is low in tar and massive in size, but too soft and fragile for use in this process. Even if it were suitable, it sells at delivered prices about equal to coal-derived coke. Fluid petroleum coke is very hard, relatively unreactive and ranges in size from 20 to 200 mesh. While it is available in some locations at as little as $6 per ton, its size and properties do not lend themselves to use in a producer gas reactor as described by Cominco. It is possible that new process design concepts might allow use of this low cost reductant either directly or by conversion to producer gas with a better chance for success than is visualized for soft coal.

2.6 Standard Flue Gases

Five different flue gas compositions were defined for the purpose of this study, per Table I below.

TABLE I

(Gas Flows reduced to 32°F, 1 atm)

Gas No.	Description	Temp. °F	Flow MSCFM	NTD S Equiv. in feed	Composition, Vol. %				
					SO_2	O_2	H_2O	CO_2	N_2
1	From Upstream Concentrator	120	6.9	400	90	-	10	-	-
2A	Undiluted Converter Gas	2300	20.0	203	15.8	1.2	.002	-	83.0
2B	Smelter Gas, Converter only	440	70.0	202	4.5	16.5	-	-	79.0
2C	Smelter Gas, Combined	440	210.0	391	2.9	14.3	0.1	1.7	80.4
3	Power Plant Stack *	300	2350.0	452	0.3	2.8	7.25	14.7	74.9

* The power plant stack gas is for a 1400 megawatt plant burning 580 NT/hr. coal containing 3.2% sulfur. The gas will also contain 0.05% NO_x and 1.15 grains of dust per SCF, the smelter gases 0.2 grains dust per SCF.

2.7 Again, in the interest of simplicity, it has been assumed that the flow rates and temperatures of these gases will be or can be made uniform at the rates given. In practice, smelter gas flows will be variable, requiring attention to scheduling to approach the kind of uniformity essential to the operation of the reduction processes described.

2.8 Considering these four reductants and the five typical gases, 20 basic combinations are possible. However, judgment in choosing the initial cases studied obviated the need for detailed study of nine of these as illustrated in Table II below, where x indicates cases studied. The time saved by this approach allowed a study in greater depth of the more important combinations and their modifications within the available time.

TABLE II

Gas No ⟶	1	2A	2B	2C	3
Coke	X	X	X	0	0
CH_4	X	X	X	0	0
H_2S	0	X	X	X	X
$H_2 + CO$	0	X	0	0	0

2.9 Each case examined entailed these 9 steps:

- Critically study the literature data
- Study the equilibria via computer
- Design a process flow sheet
- Define material and energy balances
- Specify and size major equipment
- Prepare fixed capital estimate
- Prepare operating cost estimate
- Analyze economics
- Define data deficiencies and assumptions made

2.10 Once assembled, the alternates were examined and certain processes selected for experimental study in a Phase II effort designed to make up data deficiencies, to test assumptions, and generally to firm up process costs.

3.0 Summary of Results

3.1 A total of 30 estimates were made including the basic cases above and certain modifications. These are summarized by flue gas type in Table III (foldout page 5A) and by reductant type in Table IV (foldout page 5B). Each of the cases summarized is supported by detailed operating cost estimates given in the Appendix to this report. Also in the Appendix are flow sheets, process profiles and fixed capital estimates for the basic cases. Where necessary to compare operating costs at equal sulfur capacity, the basic capital estimates prepared by Allied's Engineering Dept. were six-tenth factored as noted on the summary tables. A table of contents is furnished with the Appendix to aid in locating these details.

4.0 Limitations

4.1 Before examining the results, it is important to recognize that estimates of the type herein are useful for comparative purposes, but can in no wise be taken as absolute. Some of the more specific limitations are elaborated below.

4.2 Fixed Capital The estimates are given on a "battery limits" basis as specific plant conditions and locations are too variable to be included and still permit a valid comparison of the contending processes. As an example, the real cost of installing very large recovery facilities in an existing power plant is virtually impossible to generalize. Gas flows are so large that valves and ductwork needed to integrate the reduction plant with the boiler plant could easily outweigh the cost of process equipment.

4.3 Working Capital In a comparison of this sort, where a variety of processes yield the same product, working capital will be nearly invariant. It was therefore ignored since it would affect all investments equally.

4.4 Feed Gas Value Costs assume the SO_2-bearing feed gas has no value. Taken from a flue it may, in fact, have a negative value if chemical absorbents are otherwise required for control.

Conversely, if the gas comes from an upstream gathering process, its value can be taken at the operating cost of the collecting system. With reduction process cost separately defined, upstream debits or credits can easily be applied to obtain overall cost.

4.5 Methane and Coke Values To provide a uniform basis for study of plants in the U. S., it was necessary to use a weighted average cost at the battery limits for reductants like natural gas and coke. Referring to Exhibit 11-1, natural gas costs may vary from 15 cents to 50 cents or more per MCF. A value of 30 cents was used. Coke, even when the grade is specified (in this instance furnace grade, 3 inch size, ash fusion above 2300°F) may vary from \$21 to \$35 per ton delivered in major industrial centers. A weighted average of \$24/NT was used.

4.6 Return on Investment Total investment (which would include facilities beyond battery limits as well as working capital) has not been provided, nor has a projected sulfur price been assumed. Long-term, a selling price of \$30/long ton f.o.b. producing plant was recently suggested. There is danger in approximating this most important variable. Certainly a pegged price cannot be used with the same assurance of stability as one feels in generalizing the delivered cost of natural gas or coke, in view of past volatility of sulfur markets.

4.6.1 What is provided is an annual operating cost and a per ton sulfur product cost, with and without steam credits. In this way one can select and apply the most appropriate credits for any specific place or point in time. It is believed that this same approach is even more practical where sulfuric acid is the product.

4.6.2 The possibility of a profit in operating any of the processes examined herein appears remote. Instead, the objective, it would seem, is to find the least costly means for emission control.

4.7 Steam Credits Where large quantities of high pressure steam could be generated, part was used to power turbine-driven blowers to conserve electrical power. Because steam has a value peculiar to a given point, costs have been given with and without credit for the net steam available. High pressure steam was taken at the guideline value of 50 cents/\$MM BTU fuel equivalent.

4.8 Electricity Except for the power plant stack gas case, electricity was taken at a typical 1 cent/KWH common to chemical plants. The NAPCA guide-line figure of 6 mills/KWH is realistic at the power generating plant including sub-station expenses and was used for that case.

4.9 Combusted flue gas Claus process plants, being less than 100% efficient, evolve undesirable concentrations of H_2S. Except for the power plant case where reheat to 1000-1200°F is impractical enough natural gas has been provided in the estimates to combust fuel gas to assure zero percent H_2S in the stream leaving the SO_2 reduction plant. Disposition of the power plant exit gas is not yet resolved.

4.10 Other Cost Bases Tabular results have in all cases been expressed both annually and per NT sulfur produced. Other bases can easily be approximated from the annual costs as shown below for a 1400 megawatt power plant (2.35 MMSCFM flue gas, 580 NT/hr. coal).

Annual $/11.2 x 10^6 = mills/KWH
Annual $/2.35 =$/MMSCFM (32°F, 1 atm)
Annual $/580 x 8000 hrs =$/NT Coal Burned (3.2% S content)

For smelter gases, the following factors apply in a plant generating nominally 200 NT/day sulfur.

1.35 x Annual $/66,000 =$/NT Copper (Average ore S/Cu = 1.35)

0.580 x Annual $/66,000 =$/NT Zinc (Average ore S/Zn = 0.580)

0.276 x Annual $/66,000 =$/NT Lead (Average ore S/Pb = 0.276)

5.0 Economic Commentary An examination of the summary economic Tables III and IV and the discussions given under each process studied can be capsuled as follows.

5.1 Coke Where a free choice of reductant exists, natural gas affords a lower cost than coke. Coke may still be a contender because of a number of process virtues, viz: high yield, no catalyst or poisoning problems, ability to handle dust-laden gas and ability to carry out the process in simple, proven equipment (i.e. a typical producer gas setup). There may be locales where gas line extensions would not be economically feasible. Then coke would become the reductant of choice. The use of low cost petroleum coke is probably competitive with gas, but the process would have to be redesigned to handle it. Coal does not look usable in the processes considered.

5.2 Natural gas Methane may be used directly and is the practical precursor of reformer gas, H_2S, H_2 and CO. The studies show that methane reduction is roughly half as costly as coke. For the gases 1, 2A and 2B (and 2C by inference), the non-catalytic Asarco methane reduction (Cases 8-3, 3a and 4a) is lower in cost than the TGS, Yushkevich and Kulcsar catalytic processes, and no conjecture as to catalyst activity or life is necessary. However, the catalytic processes remain contenders because they operate at much lower temperatures, broadening the choice of materials and lengthening equipment life. If a long-lived catalyst is developed, the small cost differential may be eliminated or shifted in favor of catalytic operation.

SUMMARY OF RESULTS - REDUCTION OF SO_2 - TABLE III

Gas No.	Vol. % SO_2/O_2	Case No.	Report No.	Reductant and Process	Sulfur Produced NTD/NTY	Fixed Capital $MM	OPERATING COST $M/Yr. Gross	$/NT Sulfur Gross	Steam Cr.	Net	Sulfur Yield	Process Exit Gas, Vol. % * H_2S	SO_2	O_2	H_2O	Other	Notes
1	90/0	4	6	Coke	400/132,000	2.60	2195	16.64	0.48	16.16	99.0	0.00	0.68	0.00	16.59	82.73	
1	90/0	4	5	Coke	200/66,000	1.75	1261	19.12	0.48	18.64	99.0	0.00	0.68	0.00	16.59	82.73	Factored
1	90/0	8-3	8	Methane (Asarco)	370/122,000	2.10	1098	8.99	0.40	8.59	92.6	0.00	3.55	0.00	54.49	41.06	
1	90/0	8-1	8	CH_4 (Yushkevitch)	388/128,000	2.54	1258	9.83	0.40	9.43	96.9	0.00	1.91	0.00	56.04	42.05	
1	90/0	8-2	8	CH_4 (West-TGS)	370/122,000	2.45	1211	9.92	0.40	9.52	92.6	0.00	3.65	0.00	54.39	51.96	
1	90/0	8-4	8	CH_4 (Kulcsar)	400/132,000	2.45	1286	9.75	0.40	9.35	99.6	0.00	0.87	0.00	12.84	86.29	
1	90/0	8-3	Int.	CH_4 (Asarco)	185/61,000	1.40	691	11.34	0.40	10.94	92.6	0.00	3.55	0.00	54.49	41.06	Factored
2A	15.8/1.2	14	5	Coke	195/64,400	4.20	2105	32.69	0.50	32.19	96.1	0.00	0.48	0.00	5.40	94.02	
2A	15.8/1.2	14	7	Coke	195/64,400	4.74	2225	34.51	2.46	32.05	96.1	0.00	0.48	0.00	5.40	94.02	Heat Recovered
2A2	15.8/1.2	15	5	Coke	180/59,500	5.20	2552	42.93	0.25	42.68	89.1	0.00	0.87	0.00	25.64	83.49	Gas 2A Quenched
2A	15.8/1.2	6-1	6	H_2S in Situ	167/55,200	2.32	1385	25.18	0.00	25.18	82.2	0.00	0.42	0.00	19.97	78.61	
2A	15.8/1.2	3a	10	CH_4 (Asarco)	194/64,000	3.26	1246	19.48	1.73	17.75	95.6	0.00	0.57	0.00	20.45	78.98	
2A	15.8/1.2	3b	10	CH_4 (West)	194/64,000	3.50	1314	20.56	1.60	18.96	94.6	0.00	0.54	0.00	20.39	79.07	
2A	15.8/1.2	5a	10	Reformed CH_4-Wet	185/61,000	3.00	1253	20.54	3.12	17.42	92.0	0.00	0.79	0.00	28.84	70.37	
2A	15.8/1.2	5b	10	Reformed CH_4-Dry	190/62,700	3.00	1272	20.29	4.10	16.19	94.0	0.00	0.63	0.00	22.42	76.95	
2B	4.5/16.5	4a	10	CH_4 (Asarco)	180/59,000	9.15	3231	54.78	7.95	46.83	88.0	0.00	0.50	0.00	18.54	80.96	
2B	4.5/16.5	4b	10	CH_4 (West-TGS)	180/59,000	9.80	3463	58.74	9.15	49.59	88.0	0.00	0.50	0.00	18.54	80.96	
2B	4.5/16.5	4c	10	CH_4 (Kulcsar)	180/59,000	9.40	3354	56.87	7.25	49.62	88.0	0.00	0.50	0.00	18.54	80.96	
2B	4.5/16.5	16	5	Coke	380/125,400	20.50	12,287	97.99	15.15	82.84	93.3	0.00	0.25	0.00	3.21	96.54	Factored
2B	4.5/16.5	16	5	Coke	190/62,700	13.20	6795	108.39	15.15	93.24	93.3	0.00	0.25	0.00	3.21	96.54	Factored
2B	4.5/16.5	1	9	H_2S -Low Temp Claus	196/65,000	6.07	2799	43.07	0.85	42.22	96.9	0.00	0.11	10.78	10.93	78.18	
2B	4.5/16.5	1a	9	H_2S -Low Temp Claus	196/65,000	6.90	3222	49.55	0.00	49.55	96.9	0.00	0.11	10.78	10.93	78.18	
2B	4.5/16.5	1b	9	H_2S -Low Temp Claus	196/65,000	6.07	2729	42.02	0.00	42.02	96.9	0.00	0.11	10.78	10.93	78.18	
2B	4.5/16.5	1c	9	H_2S -Low Temp Claus	196/65,000	4.44	2437	37.52	0.00	37.52	96.0	0.00	0.10	9.23	20.62	70.05	
2B	4.5/16.5	1d	9	H_2S -Low Temp Claus	196/65,000	5.06	2674	41.16	0.00	41.16	97.6	0.00	0.09	12.69	7.32	79.90	
2C	2.9/14.3	6	10	H_2S -Low Temp Claus	385/127,000	7.70	4802	37.77	3.50	34.27	98.5	0.00	0.036	8.42	19.95	71.59	
2C	2.9/14.3	6	Int.	H_2S -Low Temp Claus	192/63,500	5.10	2707	42.58	3.50	39.08	98.5	0.00	0.036	8.42	19.95	71.59	Factored
3	0.30/2.8	2R	10	H_2S -Low Temp Claus	**385/127,000**	**16.8**	**7723**	**60.79**	2.15	**58.64**	85.0	0.03	0.015	2.78	7.92	89.26	
3	0.30/2.8	2R	Int.	H_2S -Low Temp Claus	**192/63,500**	11.5	**4535**	**71.43**	2.15	**69.28**	85.0	0.03	0.015	2.78	7.92	89.26	Factored
3	0.30/2.8	2B	Int.	H_2S -Low Temp Claus	385/127,000	19.0	6884	54.20	0.00	54.20	85.0	0.03	0.015	2.78	7.92	89.26	**Switch Reactor**
3	**0.30/2.8**	**2A**	**Int.**	**H_2S -Low Temp Claus**	**385/127,000**	**9.67**	**5575**	**46.06**	**2.15**	**43.91**	**85.0**	**0.03**	**0.015**	**2.78**	**7.92**	**89.26**	**Vert MBR's**

NOTE: Cases 8-1 thru 8-4 were designated Cases 1 thru 4 in Report No. 8, hence the prefix 8 to avoid duplication.

* After combusting residual H_2S in stoichiometric air, except gas No. 2, where volume to heat is impractical.

PH 22-68-24
GBF 6-5-69

SUMMARY OF RESULTS - LISTED BY REDUCTANT - REDUCTION OF SO_2 - TABLE IV

								OPERATING COST					Process Exit Gas, Vol. % *					
									$/NT Sulfur									
Reductant	Process	Gas No.	Case No.	Report No.	Vol. % SO_2/O_2	Sulfur Produced NTD/NTY	Capital $MM	$M/Yr. Gross	Gross	Steam Cr.	Net	Sulfur Yield	H_2S	SO_2	O_2	H_2O	Other	Notes
Coke	Lepsoe	1	4	6	90/0	400/132,000	2.60	2195	16.64	0.48	16.16	99.0	0.00	0.68	0.00	16.59	82.73	Factored
Coke	"	1	4	5	90/0	200/66,000	1.75	1261	19.12	0.48	18.64	99.0	"	0.68	"	16.59	82.73	
Coke	"	2A	14	5	15.8/1.2	195/64,400	4.20	2105	32.69	0.50	32.19	96.1	"	0.48	"	5.40	94.02	
Coke	"	2A	14	7	15.8/1.2	195/64,400	4.74	2225	34.51	2.46	32.05	96.1	"	0.48	"	5.40	94.02	Heat Recovered
Coke	"	2A2	15	5	15.8/1.2	180/59,500	5.20	2552	42.93	0.25	42.68	89.1	"	0.87	"	25.64	83.49	Gas 2A Quenched
Coke	"	2B	16	5	4.5/16.5	380/125,400	20.50	12,287	97.99	15.15	82.84	93.3	"					
Coke	"	2B	16	5	4.5/16.5	190/62,700	13.20	6795	108.39	15.15	93.24	93.3	"					
CH_4	Yushkevich	1	8-1	8	90/0	388/128,000	2.54	1258	9.83	0.40	9.43	96.9	0.00	0.25	0.00	3.21	96.54	Factored
CH_4	West-TGS	1	8-2	8	90/0	370/122,000	2.45	1211	9.92	0.40	9.52	92.6	"	0.25	0.00	3.21	96.54	Factored
CH_4	Asarco	1	8-3	8	90/0	370/122,000	2.10	1098	8.99	0.40	8.59	92.6	"	3.55	"	54.49	41.06	
CH_4	Asarco	1	8-3	Int	90/0	185/61,000	1.40	691	11.34	0.40	10.94	92.6	"	3.55	"	54.49	41.06	Factored
CH_4	Kulcsar	1	8-4	8	90/0	400/132,000	2.45	1286	9.75	0.40	9.35	99.6	"	0.87	"	12.84	86.29	
CH_4	Asarco	2A	3a	10	15.8/1.2	194/64,000	3.26	1246	19.48	1.73	17.75	95.6	"	0.57	"	20.45	78.98	
CH_4	West-TGS	2A	3b	10	15.8/1.2	194/64,000	3.50	1314	20.56	1.60	18.96	94.6	"	0.54	"	20.39	79.07	
CH_4	Asarco	2B	4a	10	4.5/16.5	180/59,000	9.15	3231	54.78	7.95	46.83	88.0	"	0.50	"	18.54	80.96	
CH_4	West-TGS	2B	4b	10	4.5/16.5	180/59,000	9.80	3463	58.74	9.15	49.59	88.0	"	0.50	"	18.54	80.96	
CH_4	Kulcsar	2B	4c	10	4.5/16.5	180/59,000	9.40	3354	56.87	7.25	49.62	88.0	"	0.50	"	18.54	80.96	
H_2S	In Situ	2A	6-1	6	15.8/1.2	167/55,200	2.32	1385	25.18	0.00	25.18	82.2	0.00	1.42	0.00	19.97	78.61	
H_2S	Low Temp Claus	2B	1	9	4.5/16.5	196/65,000	6.07	2799	43.07	.85	42.22	96.9	"	0.11	10.78	10.93	78.18	
H_2S	" " "	2B	1a	9	4.5/16.5	196/65,000	6.90	3222	49.55	0.00	49.55	96.9	"	0.11	10.78	10.93	78.18	
H_2S	" " "	2B	1b	9	4.5/16.5	196/65,000	6.07	2729	42.02	0.00	42.02	96.9	"	0.11	10.78	10.93	78.18	
H_2S	" " "	2B	1c	9	4.5/16.5	196/65,000	4.44	2437	37.52	0.00	37.52	96.0	"	0.10	9.23	20.62	70.05	
H_2S	" " "	2B	1d	9	4.5/16.5	196/65,000	5.06	2674	41.16	0.00	41.16	97.6	"	0.09	12.69	7.32	79.90	
H_2S	" " "	2C	6	10	2.9/14.3	385/127,000	7.70	4802	37.77	3.50	34.27	98.5	"	0.036	8.42	19.95	71.59	
H_2S	" " "	2C	6	Int	2.9/14.3	192/63,500	5.10	2707	42.58	3.50	39.08	98.5	"	0.036	8.42	19.95	71.57	Factored
H_2S	" " "	3	2R	10	0.30/2.8	385/127,000	16.8	7723	60.79	2.15	58.64	85.0	.03	0.015	2.78	7.92	89.26	
H_2S	" " "	3	2R	Int	0.30/2.8	192/63,500	11.5	4535	71.43	2.15	69.28	85.0	.03	0.015	2.78	7.92	89.26	Factored
Reformed	Wet Gas	2A	5a	10	15.8/1.2	185/61,000	3.00	1253	20.54	3.12	17.42	92.0	0.00	0.79	0.00	28.84	70.37	
CH_4	Dry Gas	2A	5b	10	15.8/1.2	190/62,700	3.00	1272	20.29	4.20	16.19	94.0	0.00	0.63	0.00	22.42	76.95	
H_2S	Low Temp Claus	3	2B	Int	0.30/2.8	385/127,000	19.0	6884	54.20	0.00	54.20	85.0	.03	0.015	2.78	7.92	89.26	Switch Reactors
H_2S	Low Temp Claus	3	2A	Int	0.30/2.8	385/127,000	9.67	5575	46.06	2.15	43.91	85.0	.03	0.015	2.78	7.92	89.26	Vert. MBR's

NOTE: Cases 8-1 thru 8-4 were designated Cases 1 thru 4 in Report No. 8, hence the prefix 8 to avoid duplication.

* After combusting residual H_2S in stoichiometric air, except gas No. 2, where volume to heat is impractical.

PH-22-68-24
GBF 6-5-69

5.3 Pure H_2 and Pure CO No process advantage has been found for the use of pure gases over the use of the mixture of these as found in reformer or producer gas. The pure gases will clearly be more costly than the mixture.

5.4 Reformed Methane A cost advantage is shown for reformer gas over methane reduction, and this advantage is further improved by drying the gas leaving the reformer (see Cases 5a vs 5b). Additional advantages, not yet assessed, include exothermic reduction at 1000-1100°F vs. 1800°F with methane. The lower temperatures should markedly improve equipment life, catalyst life and reduce maintenance. There are no unknowns in design or operation of reformers, so the use of reformer gas contributes to the reliability of the estimates.

5.5 Hydrogen Sulfide No cost advantage was found for H_2S as used at conventional Claus operating temperatures compared with the other reductants discussed above. A very serious disadvantage is the low yield from an H_2S process which further requires recycle of more than 2/3 the sulfur produced. The inefficiency of the process is further reflected in the higher fuel requirements to burn out the residual H_2S in the Claus exit gas (Refer to Case 6-1)

5.6 Low Temperature Claus It is concluded that the principal virtue of H_2S is its ability to react preferentially with SO_2 at low temperatures (150-300°F) in mixtures of SO_2 and O_2. Its advantage is apparent by comparing the relative costs for reducing 2B gas with methane and with H_2S. The 2B gas, of high O_2/SO_2 ratio, was as far as the comparison was carried. Clearly no reductant studied can compete with H_2S for reducing gases such as 2C and 3.

5.7 Cost effects of SO_2-O_2 concentration The cost of reductant will vary directly with the sum (SO_2+ O_2) in the feed gas, i.e. one mol of O_2 will consume the same quantity of reductant as one mol of SO_2. Thus any oxygen in the feed represents a cost penalty. Both capital and operating costs will increase as SO_2 concentration decreases, but the function is complex. Capital should increase as the 0.6 power of volume ratio, which allows comparing the same process at different capacities. But volume, besides affecting fixed charges also affects pumping costs and fuel to reheat and burn residual H_2S in the exit gas. If more data points were available for each reductant, an empirical relationship could likely be established between O_2 and SO_2 concentrations and the total operating cost. This would exemplify, if it is not already evident, the importance of avoiding air dilution if the economy of SO_2 recovery or control is to be improved.

5.8 Power Stack Gas Gas No. 3 provides a case of huge volumes at extreme dilution. The results presented are incomplete in that a cost has been estimated which provides a 90% reduction in SO_2 emission, but results in an emission containing .03% (300 ppm) H_2S which may not be acceptable. Obviously, it is not feasible to reheat the entire flow to 1000-1200°F, a practice acceptable for the lower volume cases. An acceptable process may result if an H_2S oxidation catalyst operable at or below 500°F is found. Such a catalyst would also be a benefit to the other smaller volume stack gas cases too.

5.9 Three different power stack gas cases were examined involving different processing techniques. The first of these is Case 2R, wherein the catalyst bed is moved by chain grate mechanisms, using co-flow of input gas and catalyst. This method is also adaptable to countercurrent flow. Case 2A is a cross-flow system, where the catalyst is distributed on the top of a slowly-moving bed, and drains off the bottom, much like coke feed and ash withdrawal from a producer gas unit. There may be less attrition and simpler mechanics in Case 2A than in Case 2R, and both capital and maintenance should be lower. Case 2B is one using switch reactors which are fixed bed reactors alternately operated as converter-absorbers, then regenerated, in the fashion suggested by Princeton Chemical Research (Research sponsored by NAPCA). Costs, summarized in foldouts 5A and 5B, are in brief, for 385 NTD sulfur production:

Case	Fixed Capital	Net Op. Cost/NT Sulfur
2A	$ 9.67 million	$43.91
2R	$16.8 million	$58.64
2B	$19.0 million	$54.20

At this point in time, it is believed that Case 2A provides the most reliable estimate and raises fewer questions than the other cases cited.

5.10 In Table V below, a comparison is made of the best Low Temperature Claus case operating on power stack gas with certain stack gas processes under study by others. Although the LT Claus process may appear to be at some disadvantage compared with the limestone processes (limestone taken at $2.05/ton), the latter produce no credits, they waste sulfur values and with dolomitic types produce a solid waste containing lwachable sulfates. Costs tend to equalize when sulfur is credited at $30/ton for all processes enumerated, if indeed the 70% acid from the Cat-Ox process is marketable and the projected consumption of alkalized alumina in the Bureau of Mines process is finally achieved.

TABLE V - CAPITAL AND OPERATING COSTS BEFORE CREDITS

8000 hrs./Yr. - Gas No. 3

Process		Capital Investment Total $M	$/Kw	$M/Yr.	$/NT Coal	Mills/KWH
Alkalized Alumina	(800 Mwe)	8,500	10.60	3402	1.54	0.54
Monsanto Cat-Ox	(800 Mwe)	17,000	21.35	3881	1.75	0.61
Limestone*-Dry	(1000 Mwe)**	3,950	3.95	2330	.88	0.29
Limestone-Wet	(1000 Mwe)**	8,200	8.21	2020	.76	0.28
Low Temp. Claus	(700 Mwe)	11,500	16.40	4400	1.20	0.78

* $2.05/NT limestone, 14.5% fixed charges

** Paper by A. V. Slack, Presented at NAPCA Contractors meeting, Cincinnati, Ohio, December 12, 1968.

5.11 Smelter Gases Table VI below gives the results of the smelter gases studied based on the lowest cost processes selected from foldout pages 5A and 5B. Here the results are given in terms of sulfur recovery cost per ton of metal.

Table VI

Cost of Sulfur Recovery Per Ton of Metal

(Before Sulfur Credits)

Gas——>	2A	2B	2C
Process——>	Reformed CH_4	Low Temp. Claus	Low Temp. Claus
Copper	$21.50	$49.80	$52.00
Lead	$ 4.47	$10.35	$10.80
Zinc	$ 9.39	$21.75	$22.65

After $30/NT Sulfur Credit

Copper	$19.00 credit	$ 9.30 debit	$11.50 debit
Lead	$ 3.83 credit	$ 2.35 "	$ 2.50 "
Zinc	$ 7.91 credit	$ 4.45 "	$ 5.35 "

The effect of sulfur credit is evident, as is the general effect of air dilution on operating cost.

5.12 Upstream gathering Because of the very modest cost involved in reducing Gas No. 1 (90% SO_2), it may be possible to combine an upstream gathering process with a reformed methane reduction, the latter at a cost in the range of 8 to $9/NT sulfur recovered. The combination could be less costly than a low temperature Claus operation, provided the cost of gathering did not exceed about $35/NT recovered sulfur. Study of gathering processes is, however, outside the scope of this contract.

5.13 Future Work An experimental program has been proposed to the National Air Pollution Control Administration which as outlined will:

- Study application of the Low Temperature Claus process to smelter gases. If a successful process is defined in the laboratory, field tests will be made on actual smelter gases.
- Define kinetics in a conventional Claus process under conditions actually existing where a Claus unit is the last step in a reduction process sequence.
- Determine if the Asarco process reduction is a catalytic or non-catalytic reaction.
- Investigate Kulcsar's claims of methane reduction of SO_2 using a silica gel catalyst.
- Optimize conditions for the intermediate temperature reactor used in many reduction processes.

6.0 Process Studies

This section details the technical aspects of those processes selected as worthy of further investigation. Data include equilibria, heat and material balances, thermodynamics and schematic flow sheets for the feed gases and reductants studied.

6.1.0 Technology Related to the Production of Feed Gases

We have selected five gas compositions representative of the principal emissions for use in our process studies. The following discussion will indicate the source of these gas streams so that a simple code reference number can be used in the body of this report.

6.1.1 Gas No. 1, Strong SO_2 from Upstream Collection Processes

This gas derives from three principal processes as follows and is defined as 90.0% SO_2 and 10.0% H_2O.

Cominco Absorption Process

The dilute SO_2 gas is absorbed in an alkaline ammonia solution to give ammonium sulfite. The sulfite is treated with H_2SO_4 to liberate the SO_2. The acidified solution is ordinarily stripped with air to provide feed to a H_2SO_4 plant. The sulfite can be steam stripped and rectified to give Gas No. 1. $(NH_4)_2SO_4$ is a by-product.

DMA Amine Absorption

The dilute SO_2 gas is absorbed in DMA (dimethylaniline) and the DMA is steam stripped and rectified to give Gas No. 1.

Sodium Sulfite-Zinc Oxide Process

The dilute SO_2 gas is absorbed in sodium sulfite solution and sodium bisulfite is formed. The bisulfite solution is treated with ZnO to precipitate zinc sulfite and regenerate sodium sulfite solution. The slurry is filtered and the zinc sulfite is dried and calcined to drive off SO_2 and regenerate ZnO. The strong SO_2 can be recovered as a gas of Gas No. 1 composition.

6.1.2 Gas No. 2A Undiluted Converter Gas

In copper smelting practice the ore concentrate consisting principally of iron and copper sulfides is melted and partially oxidized in a reverberatory furnace. The product of the operation is a mixture of molten copper sulfide and iron sulfide (Matte) and a molten slag which is separated.

The matte is transferred to a converter where it is blown with air to form molten copper and a fluid iron silicate slag. The molten metal (blister copper) is cast then further refined.

The gas leaving the converter is designated 2A. This hot gas is at 2300°F and carries molten slag particles. In present copper smelter practice this gas is mixed with a large amount of air to cool and solidify the slag. The large oxygen content makes the diluted gas undesirable for processing to elemental sulfur.

Smelter practice would have to be considerably modified to avoid air dilution. Tight manifolding would be required and this might be practical only in new plants designed specifically to avoid air dilution.

The gas from an individual converter is variable and a battery of converters operating on a time schedule would be necessary to give a reasonably uniform gas flow.

Copper concentrate is roasted in certain cases before reverberatory smelting. Both multiple hearth and fluid bed roasters are used. The hearth roasters give an SO_2 gas stream of 5-10% SO_2 with a large air dilution. The fluid bed roaster gives a 12-14% SO_2 gas carrying very little oxygen. This fluid bed roaster gas would be equivalent to the 2A converter gas in so far as conversion to elemental sulfur is concerned.

6.1.3 Gas No. 2B Smelter Gas

This gas is the highly air diluted copper converter gas as described under 2A above.

6.1.4 Gas No. 2C Smelter Gas

This gas represents combined dilute smelter gas streams including the gas from reverberatory furnaces and converters. It may include dilute roaster gas.

6.1.5 Gas No. 3 Power Plant Stack Gas

This gas represents the volume and composition of a typical 1400 MW Power Plant burning about 580 tons per hour of coal containing 3.2% total sulfur.

6.2.0 Cooling Undiluted Converter Gas 2A

6.2.1 The conventional practice in a copper smelter is to drop the temperature of the 2300°F converter gas to a moderate temperature by pulling in a large volume of air at the point where the gas leaves the converter. This dilution and oxygen introduction makes the air diluted gas unsuitable for direct reduction to sulfur by any of the methods where oxygen could react with the reductant. In high temperature sulfur recovery processes the reductant consumption relates directly with the sum of $SO_2 + O_2$.

6.2.2 The hot gas cannot be introduced directly into a waste heat recovery boiler due to entrained molten slag droplets. A process is visualized which might avoid oxygen entry and allow waste heat recovery from the hot gas.

6.2.3 It is proposed to cool the gas from 2300°F down to 2000°F by injecting water. A tight manifold at this point would be necessary to prevent air entry. It is assumed that at 200°F the slag will be solidified and will drop out either in the flue or in the boiler. The cooled gas is cleaned and is processed for sulfur recovery.

6.2.4 Exhibit 7-4 carries a layout for the proposed method, designating this gas as 2A1.

6.2.5 Another cooling procedure for 2A gas was considered where the gas was cooled to 900°F by water injection and cleaned in a precipitator at this temperature. This scheme is undesirable because of the high water dilution and the reduced recovery of byproduct steam. This gas is called 2A2.

6.2.6 The composition and volume resulting from the cooling of 20,000 SCFM of 2A gas by the two water injection methods is tabulated below:

Table VII

	2A	2A2	2A1
% SO_2	15.8	10.9	14.7
% O_2	1.2	0.8	1.1
% H_2O	0.0	31.1	7.2
% N_2	83.0	57.2	77.0
SCFM	20,000	29,000	21,600
Temp. °F	2,300	900	2,000
Equiv. sulfur	203	203	203

7.0 Hot Coke Reduction

Solid carbon at elevated temperatures is one of the major reductant systems which is effective in converting SO_2 to elemental sulfur. In this section are reviewed the extensive studies made on this system.

The scope of the work is to define the process as applied to several feed compositions in sufficient detail to allow realistic cost comparisons with other methods of converting SO_2 to sulfur.

7.1 Process Background

The systems proposed are based mainly on work of the Consolidated Mining and Smelting Co. of Canada, where a unit with a capacity of about 125 tons sulfur per day was operated at their Trail smelter around 1937.

This work is summarized by Robert Lepsoe in two publications and a U. S. Patent as follows:

a) U. S. Patent 2,102,000 - Dec. 14, 1937
b) "Chemistry of SO_2 Reduction" - "Thermodynamics" I & EC, 30, 92-100
c) "Chemistry of SO_2 Reduction" - "Kinetics" I & EC, 32,910-918

7.2 Brief Discussion of Thermochemistry, Chemical Equilibria and Kinetics

7.2.1 The thermochemistry has been updated and the concept of regenerative preheating developed by Allied Chemical for large scale application has been adapted. This should lead to a near optimum process for the coke reduction system.

7.2.2 The reduction step is based on a 90% reduction of SO_2 and the concurrent production of reducing compounds (CO, H_2, COS, CS_2) equivalent to the unreduced SO_2 remaining. This represents a practical balance between the exothermic reaction of SO_2 + carbon and the endothermic reaction of CO_2 + carbon to form CO.

7.2.3 The thermal calculations of Lepsoe are in considerable error due to faulty data. For the reaction:

$$SO_2 + C \longrightarrow CO_2 + 1/2\ S_2$$

Lepsoe reports a heat of reaction at 298°K (77°F) of -10,830 cal in the subject patent while the modern JANAF value is -7687 cal. Thus, the heat release is only about 71% of the value used by Lepsoe.

7.2.4 The heat of reaction of $CO_2 + C \longrightarrow 2\ CO$ used by Lepsoe was + 40,595 cal., with which the modern value is in good agreement at + 41,220 cal.

7.2.5 Dry SO_2 reacts with hot coke at elevated temperatures to form a mixture of CS_2, COS, S_2, CO and CO_2. The distribution is dependent on the relative approach to equilibrium.

7.2.6 In the presence of water, H_2 and H_2S are products of the reaction. Under equilibrium conditions H_2S predominates at low temperatures and H_2 is the principal product of the H_2O reaction at high temperatures.

7.2.7 Under the non-equilibrium conditions obtaining in the practical application of the carbon reduction the relative amounts of H_2S and H_2 is not predictable. Lepsoe and others indicate that in the presence of H_2O, large amounts of H_2S are formed. In this study, it is assumed that the hydrogen in the H_2O goes to H_2S.

7.2.8 In the temperature range of about 1550-2200°F, the reaction rate is dependent on the reactivity of the coke and is strongly dependent on the temperature. In this temperature range the rate increases rapidly with temperature but the contact time is relatively long ranging from a few seconds to several minutes. Above 2200°C, the reaction rate is diffusion controlled and is very fast, being in the fractional second range.

7.2.9 The reaction is exothermic up to about 90% SO_2 reduction. Above this point the endothermic reduction of CO_2 to CO predominates and flattens the temperature rise.

7.2.10 The design problem involved in obtaining a high reaction rate in coke reduction is to maintain the reaction zone above about 2100°F and below about 2400°F where the coke starts to soften and become sticky.

7.2.11 The practical solution of this problem has been defined by Lepsoe of Consolidated Mining and Smelting Co. and demonstrated by large-scale use. The process is based on the introduction of SO_2 at several levels in the reduction unit to maintain a wide zone in the high temperature range. The quantity and degree of preheat may be varied to control the temperature profile in the unit.

7.2.12 The product gas from the coke reduction is cleaned of dust by electrical precipitation or other methods and the SO_2 content is adjusted to react CS_2, COS, CO and H_2 to CO_2, H_2O and H_2S over a catalyst at about 850°F. Sulfur is recovered from this gas in a conventional Claus unit.

7.2.13 The key features of the hot coke reduction system may be summarized as follows:

a) It is non-catalytic. It is not affected by over-temperature, dust or catalyst poisons.
b) Coke is a commonly available reductant.
c) Temperature can be regulated by controlled air addition.
d) Equipment is simple. The reactor is essentially a modified gas producer.

7.2.14 The initial phase of this study of the hot coke reduction system was a series of calculations necessary to define the optimum feed conditions to give a thermally self-sufficient system. The regenerative feed gas preheating system proposed requires an adiabatic heat rise in the reduction unit of about 480°F. These initial calculations are primarily to define feed conditions which will give the required heat rise with a minimum consumption of coke as fuel. The 480°F adiabatic heat rise includes an allowance for a 7% heat loss so the actual heat rise in an operating coke reduction unit would be less than 480°F.

7.2.15 At the reaction temperature used, Lepsoe found CO to be the only reducing gas formed. CS_2 would be predicted, but apparently the reaction forming CS_2 is slow in comparison with the reduction. COS was found but was attributed entirely to the reaction of CO and S_2 which took place during the cooling of the gas samples.

7.2.16 In the above system S_2 is the only sulfur species present. It is postulated that the heat rise might be increased by the addition of H_2O. Our basis for this derives from the fact that S_2 is a very endothermic species and carries a large energy content. Any other sulfur compound which might exist under reaction conditions would carry less energy and result in a larger temperature rise. The hydrogen from the water allows H_2S as a possible component.

7.2.17 Concern on heat rise resulted from preliminary calculations on 100% SO_2 which showed a temperature drop and indicated the need for supplemental air or oxygen to arrive at the required condition. Any increase in heat rise by H_2O addition would decrease the oxygen requirement and reduce the carbon consumption factor.

7.2.18 The heat rise calculations are summarized in Exhibit 4-2. The heat rise vs. the inlet and product compositions are of principal concern. The following comments are to clarify the reason for calculating particular cases. A heat rise of 480°F is the target.

Case 1 - Defines the need for additional exothermic effect.

Case 2 and 3 together with 4 and 5 compare the effect of the hydrogen in the H_2O going either to H_2 or H_2S. Both cases improve the heat rise but H_2S as a product is most favorable.

Case 6 - oxygen addition

Case 7 - (90 SO_2 + 10 H_2O) + 25% air (gas #1)

Case 8 - (90 SO_2 + 10 H_2O) + 50% air (gas #1)

Case 9 - SO_2 + CH_4

Case 10 - SO_2 + H_2O + O_2

Case 11 - Smelter gas (gas 2C)

Case 12 - Converter gas (gas 2A)

Case 13 - Converter gas + water (gas 2A)

Case 14 - Converter gas 2A cooled indirectly and with added water and air.

Case 15 - Converter gas 2A2, i.e., converter gas 2A cooled from 2300°F to 900°F by direct water injection. The gas feed composition differs from standard Gas 2A2 since additional air was needed to supply additional heat in the reduction step.

Case 16 - Smelter gas 2B. Part of the gas reduced on carbon with the generation of a large amount of CO to limit the temperature rise in the reduction unit due to the high oxygen. Balance of the smelter gas to be added in a combustion chamber and the temperature reduced in a waste heat boiler.

7.2.19 The conclusions drawn from these preliminary heat balance studies were as follows:

a) The use of a proper H_2O content will minimize or eliminate the need for supplemental oxygen on strong SO_2 or on low oxygen converter gas 2A.

b) Direct processing of smelter gas 2B or 2C leads to excessive temperatures and high carbon consumption.

c) Case 16 represents a possible process approach to smelter gas. Feed gas is at a low temperature and the temperature rise is limited to an acceptable maximum by operating to allow the endothermic formation of CO to go to a large degree. The excess CO would be utilized by adding supplemental smelter gas in a combustion chamber and the temperature reduced in a waste heat boiler. Extensive kinetic information would be required to allow a detailed appraisal of this approach.

d) The distribution of the hydrogen from added H_2O between H_2 and H_2S cannot be estimated. Under conditions approaching equilibrium this could be done, but this very endothermic reaction would cause the temperature to drop and the system would be inoperable. Equilibrium studies were not conducted on the carbon reduction system. This process operates so far from equilibrium that such studies would have no significance.

7.3 Coke Reduction Process Studies

7.3.1 Four detailed process studies were made and were subjected to a comprehensive evaluation. These studies covered conditions as follows.

Case 4 - Coke reduction of No. 1 gas with added steam to maximize heat release. Flows are based on 500 tons sulfur production per day. Economic studies were scaled to 200 and 400 ton per day of sulfur for comparative purposes. Refer to Exhibits 5-2 and 5-3.

Case 14 - Coke reduction of 2A gas with added steam and air. It is based on the indirect cooling of the 2300°F gas by some unspecified process. Refer to Exhibits 5-4 and 5-5.

Case 15 - Coke reduction of 2A2 gas with added air for heat balance. It is based on the cooling of the 2300°F 2A gas by water injection. Refer to Exhibits 5-6 and 5-7.

Case 16 - This is based on coke reduction of part of the 2B smelter gas under conditions giving a large formation of CO as a means of limiting the temperature rise to an allowable degree. This gas containing excess CO is proportioned with additional high oxygen 2B gas and burned in a combustion chamber to give a gas composition proper for the intermediate converter. Refer to Exhibits 5-8 and 5-9.

7.3.2 The process parameters are set to conform with the following considerations:

a) Use of controlled H_2O addition to maximize the heat release in the reduction.

b) Use of additional air to augment the heat release where required by heat balance considerations.

c) Full use of regenerative preheating of the feed gas to minimize supplemental air.

d) A 7.0% heat loss based on the enthalpy at the maximum gas temperature is assumed.

e) A 2200°F temperature leaving the reduction unit and a maximum hot zone temperature of 2400°F to prevent agglomeration of the coke bed.

f) The nature of the process following the reduction unit is varied to suit the particular properties of the gas in the several cases.

7.4 Factors for Scale-up and Equipment Sizing

Pertinent factors in the scale-up and equipment sizing are summarized as follows.

7.4.1 The Reduction Unit The reduction unit is scaled from Lepsoe with modifications as indicated in the following.

a) Standard gas producers operate on egg coke (3"-2-1/2") at a firing rate of 65-70#/hr./sq.ft. grate area.

b) Lepsoe shows SO_2 reduction on a unit operating at 64# coke/hr/sq.ft. in which the reaction is complete to SO_2 equilibrium in 20" of coke bed. The reaction is complete to a ratio of $\frac{CO}{SO_2}$ = 2 in 10" of bed. The total bed depth was 50". Coke was 3/8-5/8" size.

c) Lepsoe shows kinetics which indicate that the rate of reduction of SO_2 on coke is 5-10 times faster than the reduction of CO_2.

7.4.2 The proposed reduction unit is rated at 125# coke/sq. ft. per hr. for the following reasons:

a) The practical introduction of supplemental SO_2 feeds requires a larger depth to diameter ratio than the regular producer gas unit in which the diameter greatly exceeds the bed depth.

b) Very large diameter units with shallow beds would probably be impractical due to difficulties in ash removal and in maintaining uniform gas flow distribution. The projected 13 ft. diameter unit is regarded as the largest practical diameter.

7.4.3 A bed depth of 12' was arbitrarily assumed. The diameter probably would be constricted at the points of supplemental feed. If this geometry is impractical, a further increase in bed depth and a larger coke size is suggested, rather than going to multiple units.

7.4.4 Egg coke is specified for this operation. It should be low in organic volatile matter (tar, etc.) since this will contaminate the sulfur.

7.4.5 The chemical reactivity of the coke is not important since at the reaction temperature the reaction is diffusion-controlled.

7.4.6 The limiting factor in the reduction unit is taken as either a coke burning rate of 125 lb./sq.ft./hr. or a superficial gas velocity of 7.5 ft./sec. at 2200°F.

7.4.7 The burning rate is limiting in Case 4 and the gas velocity is limiting in Cases 14, 15 and 16. The number of units projected for Case 16 may be too few if the gas flow must be considerably slowed to allow the projected production of CO.

7.5 The Regenerator System This system is scaled from detailed engineering studies made by Allied for similar heat regeneration applications.

7.6 The Electrostatic Precipitator A gas entry temperature of 840°F max. is projected and a gas cooler is provided where necessary.

7.7 The Intermediate Catalyst Converter A 2.0 sec. residence time is provided. This unit is not required when the reducing gas is almost entirely H_2S. The H_2 and CO content is not predictable and it seemed realistic to provide the converter in these projections. Laboratory work is necessary to resolve this point specifically.

7.8 The Claus Unit The catalyst is distributed in two beds with a residence time of 2.0 sec. in each. Interstage sulfur removal is incorporated in this projection.

The size and number of each of the process units is summarized in Exhibit 5-10 for the four cases under study.

Commercial Plant Projection

7.9 Exhibit 5-10 summarizes the sizing and other pertinent data on these four cases. The information in this exhibit, together with the flow sheets and process profiles, provided the basis for the fixed capital and utilities estimates made by Allied's Engineering Department.

8.0 Reduction of Sulfur Dioxide with Methane

8.1 Background A considerable amount of information is available in the literature pertaining to the reduction of sulfur dioxide with methane. The principal products of this reduction are elemental sulfur, hydrogen sulfide, water and carbon dioxide. The by-products are hydrogen, carbon monoxide and small amounts of carbonyl sulfide. The elemental sulfur is present as S_2 at temperatures above 1600°F but increasing proportions of the polymerized sulfur species, S_6 and S_8, exist at lower temperatures.

8.2 Primary Reactions

8.2.1 A casual survey of the literature data might cause one to thing that information is inadequate and contradictory. A close examination of the data reveals consistent patterns leading to logical conclusions. The two principal reactions can be represented by:

$$(1) \quad 2\ SO_2 + CH_4 \longrightarrow S_2 + CO_2 + 2\ H_2O$$

$$(2) \quad 4\ SO_2 + 3\ CH_4 \longrightarrow 4\ H_2S + 3\ CO_2 + 2\ H_2O$$

Since the equilibrium constants of reactions (1) and (2) are high (See Table VIII) over the normal range of reduction temperature, the reactions should go practically to completion. In fact, if one considers the several possible reactions in which methane reacts (See Exhibit No. 11-2) it is apparent that the methane should be completely consumed under equilibrium conditions at temperatures above 1400-1500°F.

TABLE VIII

Temperature			Log Kp *		Kp	
°K	°C	°F	Reaction(1)	Reaction(2)	Reaction(1)	Reaction(2)
800	526.84	980.31	12.612	37.618	4.09×10^{12}	4.15×10^{37}
900	626.84	1160.31	12.065	35.289	1.16×10^{12}	1.94×10^{35}
1000	726.84	1340.31	11.625	33.421	4.22×10^{11}	2.64×10^{33}
1100	826.84	1520.31	11.260	31.888	1.82×10^{11}	7.73×10^{31}
1200	926.84	1700.31	10.956	30.614	9.04×10^{10}	4.11×10^{30}
1300	1026.84	1880.31	10.691	29.525	4.91×10^{10}	3.35×10^{29}
1400	1126.84	2060.31	10.463	28.593	2.90×10^{10}	3.92×10^{28}
1500	1226.84	2240.31	10.263	27.779	1.83×10^{10}	6.01×10^{27}
1600	1326.84	2420.31	10.085	27.065	1.22×10^{10}	1.16×10^{27}
1700	1426.84	2600.31	9.929	26.437	8.49×10^{9}	2.73×10^{26}
1800	1526.84	2780.31	9.785	25.873	6.10×10^{9}	7.46×10^{25}

* Data for each molecular species from JANAF tables

Reaction(1) = Reaction (A), Exh. 11-2

These results are shown graphically in Exhibits 8-1 through 8-8.

8.2.2 In view of this fact, Allied calculated equilibrium compositions (see computer write up) for numerous feed compositions for the methane reduction of sulfur dioxide as reported in the literature. The equilibrium product gases at anticipated reaction temperature contained no unreacted methane, a negligible amount of carbon disulfide (CS_2) and only small amounts of carbonyl sulfide (COS). This is exemplified by the equilibrium products resulting from reactions (1) and (2) as shown in Table IX and Exhibits 11-3 and 8-6 appended.

TABLE IX

Equilibrium Reaction Products at Various Temperatures For the Reduction of Sulfur Dioxide with Methane

SO_2/CH_4 Feed Ratio	Product Ratios	Temperatures, °F 1000	1200	1400	1600	1800	2000	2200	2400
1.33	SO_2/ S	.0001	.0002	.0006	.0014	.0033	.0070	.0138	.0246
"	H_2S/ S	.976	.936	.868	.763	.638	.509	.396	.303
"	$2S_2$/ S	.0136	.0437	.107	.208	.334	.462	.575	.662
"	$6S_6$/ S	.0000	-	-	-	-	-	-	-
"	$8S_8$/ S	.0000	-	-	-	-	-	-	-
"	COS/ S	.0117	.0194	.0255	.0272	.0247	.0203	.0158	.0120
"	CO_2/ C	.980	.951	.896	.813	.716	.618	.531	.456
"	CO/ C	.0037	.0231	.0696	.149	.250	.353	.447	.527
"	H_2O/ H_2	.344	.358	.385	.425	.471	.514	.549	.573
"	H_2/ H_2	.0068	.0180	.0378	.0671	.104	.145	.187	.226
2.00	SO_2/ S	.162	.155	.139	.131	.131	.141	.159	.183
"	H_2S/ S	.321	.305	.264	.229	.196	.165	.136	.112
"	$2S_2$/ S	.341	.534	.593	.636	.667	.688	.700	.701
"	$6S_6$/ S	.0328	.0012	.0002	.0000	-	-	-	-
"	$8S_8$/ S	.141	.0006	.0000	-	-	-	-	-
"	COS/ S	.0020	.0034	.0045	.0054	.0060	.0060	.0059	.0050
"	CO_2/ C	.995	.991	.983	.966	.929	.871	.793	.705
"	CO/ C	.0002	.0016	.0072	.0233	.0582	.116	.195	.285
"	H_2O/ H_2	.678	.693	.729	.756	.773	.783	.786	.780
"	H_2/ H_2	.0006	.0023	.0067	.0157	.0307	.0517	.0779	.107

Note: CH_4 and CS_2 less than 0.0000. See Exhibits 11-3 and 11-4.

8.2.3 The equilibrium conversions were verified by the computations of Averbukh in Table X below:

TABLE X

SO_2/CH_4 Feed Ratio	Temperature °K	°F	S_2	H_2S	CS_2	COS	SO_2	Reference
2.0	1000	1340	58.8	26.6	-	0.6	14.0	Averbukh (4)
"	"	"	58.2	27.5	0.0	0.0	14.3	Allied
"	1100	1520	62.8	23.5	-	0.7	13.0	Averbukh (4)
"	"	"	62.4	24.3	-	0.0	13.3	Allied
"	1300	1880	68.2	17.8	-	0.8	13.2	Averbukh (4)
"	"	"	68.3	18.3	-	0.0	13.4	Allied
"	1500	2240	70.4	12.8	-	0.7	16.1	Averbukh (4)
"	"	"	70.7	13.0	-	0.0	16.3	Allied
1.33	1000	1340	16.1	79.9	0.1	3.8	0.1	Averbukh (4)
"	"	"	8.4	89.2	0	2.4	0.0	Allied
"	1300	1880	38.5	58.0	-	3.1	0.4	Averbukh (8)
"	"	"	39.0	58.7	-	2.3	0.0	Allied
"	1500	2240	59.4	37.2	-	1.9	1.5	Averbukh (4)
"	"	"	59.4	37.7	-	1.5	1.4	Allied

8.2.4 From these results one can see that slight changes of SO_2/CH_4 feed ratio between 1.33 and 2.0 can cause significant changes in product composition, especially in regard to the amount of elemental sulfur, H_2S, and unreacted SO_2. Furthermore, it should be noted that the maximum conversion of SO_2 to elemental sulfur at a $2/1:SO_2/CH_4$ feed ratio is only slightly more than 70%.

8.2.5 The literature data was compared with equilibrium data with excellent agreement when high reaction temperatures and good catalysts were used at reasonable space velocities:

TABLE XI

Component	West(1) Data	Calc. Equil. Comp.	Yushkevich(2) Data	Calc. Equil. Comp.	Kulcsar SVH=148	Data(3) SVH=548	Calc. Equil. Comp.
S_2	16.9	20.4	16.18	17.39	23.82	15.70	20.98
H_2S	15.1	12.6	11.52	9.71	3.65	2.69	8.47
COS	3.7	0.0	0.31	0.24	0.09	2.89	0.20
SO_2	9.4	8.6	7.02	6.38	20.82	34.97	21.57
H_2O	53.8	54.1	38.45	38.77	49.68	29.32	46.40
CO_2	27.3	28.6	24.32	23.74	27.90	20.87	27.05
CO	0.0	2.0	0.97	1.19	0.17	0.53	0.54
H_2	0.0	2.4	1.23	1.58	0	16.79	0.73
CH_4	Nil	Nil	Nil	Nil	1.07	3.49	0.00
CS_2	"	"	"	"	0.0	0.12	0.00

(1) Texas Gulf Sulfur's U. S. Patent 3,199,965. Initial feed composition of 62.06 moles SO_2, 31.03 moles CH_4, and 6.91 moles of H_2O.

(2) Ref. No. 369 , Exp. 5 and 6, 1st reactor at 1650°F and SVH=500 in series with 2nd reactor at 700°F, SVH=240.

(3) Ref. No. 108 , 1650°F Feed ratio of SO_2/CH_4 = 2.6. This data shows that sulfur yields of 60-80% are attainable at 1650-1830°F. For this reason additional lower temperature reactors are required to obtain overall sulfur yields above 90%.

8.2.6 Kulcsar's data at an SVH of 584 was included to show the change in product composition and the corresponding deviation from equilibrium merely by increasing the SVH from 148 to 584.

8.2.7 After concluding that equilibrium can be easily attained by the use of high temperatures, low to moderate space velocities and/or the use of good catalysts (activated alumina, silica gel, activated bauxite or Allied's proprietary special catalyst, the literature was reviewed to see whether or not the apparent discrepancies in the reported data could be rationalized. In every instance, the departure from equilibrium could be explained by the use of too low a temperature, the use of a less active catalyst (quartz) or too high a space velocity for the temperatures and catalyst (if any) employed. The data of Kulcsar and Averbukh, Exhibits 11-5a and 11-5b, in particular illustrate this fact. Large amounts of unreacted methane and sulfur dioxide, as well as substantial quantities of hydrogen, carbon disulfide and/or carbonyl sulfide are usually present in the gaseous products under certain conditions that are unfavorable for equilibrium as cited above.

8.2.8 Some researchers have indicated that carbon may be formed which could impair the quality of the product sulfur. This carbon is apparently produced under conditions favoring poor combustion (i.e. far from the equilibrium situation). Carbon will probably not be produced in the process profiles employing an active catalyst nor in the process profiles operating without a catalyst at high temperatures, providing the space velocities are not too high.

8.3 PROCESS PROFILES

8.3.1 In the process evaluations on methane reduction, conditions were assumed favoring equilibrium and accordingly equilibrium data are used except for a few instances as noted later. Since the Asarco and Texas Gulf Sulfur (West) processes, as well as the Yushkevich laboratory work, used a three step process to obtain high overall conversions of SO_2 to sulfur (usually 90% or higher), the same principle was followed, i.e.

- A high temperature reduction step to reduce the SO_2 to elemental sulfur, S_2, with the concurrent conversion of the methane to carbon-containing products, CO_2, CO, and COS, as well as hydrogen-containing substances, H_2S, H_2O, and H_2.
- An intermediate temperature conversion step for conversion of the by-product H_2, CO and COS to S_2, CO_2 and H_2O.
- A low temperature Claus-type reactor (or reactors) for reacting SO_2 and H_2S to form additional sulfur.

8.3.2 Four different high temperature reduction methods were selected from the literature as bases for the process profiles. These four schemes were selected since they portray four moderately different techniques for achieving equilibrium:

Process	Reaction Temperature °C	Reaction Temperature °F	Space Velocity (SVH *)	Catalyst
Yushkevich	900	1650	350	Bauxite
Texas Gulf Sulfur (West)	1000	1830	500	Activated Alumina
Kulcsar	1100	2010	6500(1st) 500(2nd)	Silica Gel
Asarco	1250	2200	1000	None

* Volume of feed gas at 32°F per volume of empty reactor per hour.

8.3.3 Since complete processes were not defined, practical composite processes were devised for the standard flue gases, as previously stated.

8.4 PROCESS CHEMISTRY

8.4.1 High temperature reduction of SO_2 with CH_4 proceeds at 1470-1830°F with catalyst (activated alumina, silica gel, bauxite, etc.), or at temperatures above 2200°F (Asarco process) without a catalyst. The principal sulfur-containing end-products are S_2 and H_2S. These reactions can be represented by:

$$2\ SO_2 + CH_4 \longrightarrow S_2 + CO_2 + 2\ H_2O$$

$$SO_2 + CH_4 \longrightarrow H_2S + CO_2 + H_2$$

A secondary reaction, such as:

$$CO_2 + H_2 \longrightarrow CO + H_2O$$

could explain the formation of by-product CO. CO and S_2 react to form COS:

$$2\ CO + S_2 \longrightarrow 2\ COS$$

8.4.2 For the overall process, the first reaction can be assumed to take place within the three step reaction system (high temperature reactors, intermediate converter, and normal temperature Claus) since the input feeds are SO_2 and CH_4, while the main components of the plant exit streams are sulfur, carbon dioxide and water. It is for this reason that the input system feed ratio of SO_2/CH_4 is 2/1 (Henceforth the term "feed ratio" applies to the proportion of SO_2 to CH_4 in a reactor feed).

8.4.3 At lower temperatures (in the intermediate converter), it is assumed that H_2, CO, and COS react completely. This assumption applies throughout the process assessments. Reactions that might occur are:

$$2\ H_2 + S_2 \longrightarrow 2\ H_2S$$

$$2\ CO + S_2 \longrightarrow 2\ COS$$

$$2\ COS + SO_2 \longrightarrow 2\ CO_2 + 3/2\ S_2$$

At equilibrium and reasonable feed rates, the amount of COS produced appears negligible. At high feed rates and low reaction temperatures, a non-equilibrium condition apparently exists and an increased proportion of COS may be formed.

8.4.4 Evidence in support of this conclusion based on Kulcsar's work is presented below:

TABLE XII

Reaction Temperature (°F)	Feed(1) Ratio	SVH(2)	Moles of COS		Total Moles of Product Gas	
			At Equil.	Observed	At Equil.	Observed
1652	2.6	148	0.20	0.09	125.94	127.20
1832	2.6	148	0.22	0.11	126.55	125.26
2012	3.1	6674	0.16	0.57	124.01	121.21
1832	2.6	584	0.22	0.92	126.55	130.60
1832	3.1	6674	0.15	0.81	123.49	124.83
1652	2.6	584	0.20	2.89	125.94	127.20

(1) Feed Gas = 100 moles of feed gas, SO_2 and CH_4, with the ratio of SO_2/CH_4 as indicated; i.e., for a feed ratio of 3.1 the feed gas contains 75.60 moles of SO_2 and 24.40 moles of CH_4.

(2) Volumes of feed gas measured at 32°F per empty reactor volume (neglects volume of catalyst) per hour.

8.4.5 The reaction that takes place in Claus units is:

$$2\ H_2S + SO_2 \longrightarrow 3/2\ S_2 + 2\ H_2O$$

8.4.6 The previous reactions portray fairly accurately the chemical reactions taking place in each part of the system, however, the equilibrium constants were based on eight non-restrictive equations,

1. $4\ H_2O + S_2 \longrightarrow 4\ H_2 + 2\ SO_2$
2. $CO + 3\ H_2 \longrightarrow H_2O + CH_4$
3. $2\ H_2 + S_2 \longrightarrow 2\ H_2S$
4. $CO + H_2O \longrightarrow H_2 + CO_2$
5. $2\ CO + S_2 \longrightarrow 2\ COS$
6. $S_2 + CO + H_2 \longrightarrow H_2O + CS_2$
7. $3\ S_2 \longrightarrow S_6$
8. $4\ S_2 \longrightarrow S_8$

The computer output data for this set of reactions is given in Table XIII overleaf.

TABLE XIII

Values of equilibrium constants at various temperatures and one atmosphere pressure.

Reaction Constant	Temperature (°F) 600.000	700.000	800.000	900.000	1000	1100	1200	1300	1400	1500	1600	1700
K1	0.2892 E-19	0.2577 E-17	0.1161 E-15	0.3066 E-14	0.5275 E-13	0.6409 E-12	0.5846 E-11	0.4197 E-10	0.2462 E-09	0.1217 E-08	0.5186 E-08	0.1946 E-07
K2	0.4771 E+07	0.1015 E+06	0.3894 E+04	0.2377 E+03	0.2103 E+02	0.2516 E+01	0.3859 E+00	0.7284 E-01	0.1632 E-01	0.4255 E-02	0.1261 E-02	0.4176 E-03
K3	0.6275 E+11	0.2822 E+10	0.2047 E+09	0.2161 E+08	0.3083 E+07	0.5615 E+06	0.1251 E+06	0.3294 E+05	0.9992 E+04	0.3418 E+04	0.1296 E+04	0.5371 E+03
K4	0.3308 E+02	0.1670 E+02	0.9484 E+01	0.5898 E+01	0.3940 E+01	0.2787 E+01	0.2065 E+01	0.1590 E+01	0.1264 E+01	0.1032 E+01	0.8617 E+00	0.7336 E+00
K5	0.1014 E+09	0.3924 E+07	0.2558 E+06	0.2507 E+05	0.3394 E+04	0.5969 E+03	0.1300 E+03	0.3381 E+02	0.1020 E+02	0.3491 E+01	0.1330 E+01	0.5559 E+00
K6	0.7052 E+06	0.5411 E+05	0.6218 E+04	0.9796 E+03	0.1985 E+03	0.4929 E+02	0.1447 E+02	0.4879 E+01	0.1849 E+01	0.7736 E+00	0.3323 E+00	0.1726 E+00
K7	0.9821 E+07	0.8139 E+05	0.1590 E+04	0.5875 E+02	0.3536 E+01	0.3119 E+00	0.3733 E-01	0.5728 E-02	0.1079 E-02	0.2414 E-03	0.6246 E-04	0.1831 E-04
K8	0.8382 E+12	0.6637 E+09	0.1679 E+07	0.1048 E+05	0.1338 E+03	0.3040 E+01	0.1106 E+00	0.5946 E-02	0.4427 E-03	0.4343 E-04	0.5383 E-05	0.8172 E-06

TABLE XIII (Continued)

Reaction Constant	1800	1900	2000	2100	2200	2300	2400	2500
K1	0.6531 E-07	0.1988 E-06	0.5554 E-06	0.1437 E-05	0.3474 E-05	0.7902 E-05	0.1701 E-04	0.3487 E-04
K2	0.1523 E-03	0.6048 E-04	0.2587 E-04	0.1182 E-04	0.5732 E-05	0.2928 E-05	0.1569 E-05	
K3	0.2405 E+03	0.1152 E+03	0.5861 E+02	0.3142 E+02	0.1765 E+02	0.1034 E+02	0.6291 E+01	0.3958 E+01
K4	0.6349 E+00	0.5574 E+00	0.4956 E+00	0.4454 E+00	0.4041 E+00	0.3798 E+00	0.3410 E+00	0.3165 E+00
K5	0.2517 E+00	0.1222 E+00	0.6306 E-01	0.3485 E-01	0.1963 E-01	0.1171 E-01	0.7251 E-02	0.4648 E-02
K6	0.9009 E-01	0.4970 E-01	0.2879 E-01	0.1741 E-01	0.1094 E-01	0.7111 E-02	0.4766 E-02	0.3284 E-02
K7	0.5982 E-05	0.2078 E-05	0.8166 E-06	0.3458 E-06	0.1564 E-06	0.7506 E-07	0.3795 E-07	0.2011 E-07
K8	0.1477 E-06	0.3110 E-07	0.7482 E-08	0.2024 E-08	0.6075 E-09	0.2000 E-09	0.7151 E-10	0.2753 E-10

8.4.7 The subsequent studies on the ASARCO (Fleming and Fitt) process, indicated that less heat was evolved at the higher temperature than at the lower temperatures used by West, Yushkevich, and Kulcsar. This is due to the greater production of CO and H_2. A table (XIV) of heats of reaction at 1880°F and 77°F for some of the many reactions which could take place in these systems is presented to better portray this point. The heat of reaction of SO_2 and H_2S at 440°F is shown depicting the formation of the molecular species $S_{7.45}$. Heat of reaction is expressed as K Cal/g mol of the underlined feed material.

TABLE XIV

Reaction	Underlined Feed Material Heat of Reaction, Kcal/g mol	
	at 1880°F	at 77°F
$3H_2 + \underline{SO_2} \longrightarrow H_2S + 2H_2O$	-54.829	-49.529
$2CO + \underline{SO_2} \longrightarrow 2CO_2 + 1/2\ S_2$	-48.050	-48.907
$\underline{2SO_2} + 4H_2 \longrightarrow S_2 + 4H_2O$	-32.816	-29.229
$\underline{H_2} + 1/2\ S_2 \longrightarrow H_2S$	-21.653	-20.300
$\underline{CO} + 1/2\ S_2 \longrightarrow COS$	-20.239	-22.083
$\underline{4SO_2} + 3CH_4 \longrightarrow 4H_2S + 3CO_2 + 2H_2O$	-19.392	-19.951
$\underline{6SO_2} + 4CH_4 \longrightarrow S_2 + 4COS + 8H_2O$	-10.717	-11.100
$\underline{2SO_2} + CH_4 \longrightarrow CO_2 + 2H_2O + S_2$	- 9.431	- 9.511
$\underline{SO_2} + 2COS \longrightarrow 3/2\ S_2 + 2CO_2$	- 5.572	- 4.741
$\underline{SO_2} + 2H_2S \longrightarrow 3/2\ S_2 + 2H_2O$	+10.490	+11.371
$\underline{CH_4} + 2H_2O \longrightarrow CO_2 + 4H_2$	+46.770	+39.437
$\underline{CH_4} + 2H_2O \longrightarrow CO + 3H_2$	+54.387	+49.276
At 227°C		
$\underline{SO_2} + 2H_2S \longrightarrow 0.403\ S_{7.45} + 2H_2O$	-22.468	

8.5 Yushkevich - Reduction of Gas No. 1 (90% SO_2 - 10% H_2O)

8.5.1 The experimental work for this process was conducted over a temperature range of 1290-1830°F employing a bauxite catalyst. The results of the experiments show that at 1650°F and 1830°F methane reacts completely at a space velocity of 500, but not at an SVH of 1000. In the latter case, the gas mixture after reaction contains 2.1% and 0.7% CH_4 at 1650°F and 1830°F respectively. 1650°F was selected as the lowest practical temperature for methane reduction. The process profile (See Exhibit 8-1) and flow sheet (See Exhibit 8-2) were based on experiments 5 and 6 of this study which took place under equilibrium conditions at 1650°F and an SVH of 350.

8.5.2 This initial process visualization embodied several assumptions to simplify computations since computer equilibrium data was not available at the time.

- Product composition was based on experimental rather than equilibrium values.
- 100% conversion of the H_2, CO, and COS was assumed to take place in the intermediate converter.
- 75% conversion of the H_2S and SO_2 was assumed for the Claus reactors.
- Sulfur was assumed to exist as S_2 although at temperatures below 1470°F, it is realized that sulfur will be present as other molecular species (usually a mixture of S_2, S_6 and S_8, typically represented as S_x).
- Heat balance computations assumed a 7% heat loss based on the enthalpy of the products at the selected reaction temperatures, i.e. the heat loss at 1470°F =

  (Enthalpy at 1470°F) x 0.07

8.5.3 The following process description applies to the Yushkevich reduction process. This description is relevant to the other methane reduction methods except for a few modifications that will be described wherever such changes occur.

8.5.4 Typical equipment for methane reduction of SO_2 includes:

- One or two High Temperature Reactors
- Two Fixed Bed Heat Regenerators
- An Intermediate Converter
- A Sulfur Condenser
- Two or more Claus Converters with a Sulfur Condenser following each unit
- A Coalescer
- An Incinerator and Stack
- Auxiliary Heat Exchangers

8.5.5 87% of the strong SO_2 feed gas (Gas No. 1) and 50% of the methane feed are forced by the feed gas blower via the flow reversing valve through one of two refractory-packed regenerator towers. The feed gas mixture preheated from 100°F to 1290°F, flows into the first of two bauxite-packed, high-temperature reactors.

8.5.6 The heat of reaction increases the temperature of the reactants from 1290°F to 1680°F (ca. 900°C), the desired reaction temperature. The hot residue gases are cooled by admixture with the remaining SO_2 and CH_4 feed gases to 1310°F before entering the second reactor. The product gases exit at 1610°F and enter the second regenerator, thereby cooling this gas while heating the refractory packing.

8.5.7 The regenerators are cyclic in operation, the refractory packing in each unit being alternately heated by the hot reaction gases and then cooled by the entering feed gases during the preheat cycle. The two reactors are required since the heat of reaction would raise the temperature about 900°F in a single reaction stage. It is conceivable that a single reactor might be employed via the use of internal or interstage cooling, but the design of such a unit depends on a precise knowledge of reactor temperature profiles and kinetic data for the range of feed conditions that might be employed. The dual reactor unit features simplicity of design and flexibility of operation with variable feeds.

8.5.8 The product gases leaving the regenerator are further cooled to 730°F before passing through the intermediate converter, packed with activated alumina. Further reactions occur and the product gas is essentially a mixture of H_2S, SO_2, CO_2 and H_2O.

8.5.9 The sulfur is condensed out of the gas stream by the primary condenser. In accord with the literature a sulfur condensation temperature of 248°F-250°F is shown, though this may be too low for practical operation since the freezing point of sulfur is approximately 239.4°F. A preferred safe operating temperature range would be 275-295°F since between 248°F and ca. 318°F liquid sulfur becomes more fluid with increasing temperature. Sulfur exhibits an unusual property in that it becomes very viscous between 318.5°F and about 570°F, so temperatures between these limits must be avoided in handling liquid sulfur. The viscosity reaches a maximum at about 370°F and then decreases (the fluidity increases) steadily to the boiling point of sulfur (832.3°F).

8.5.10 Since the high water content of the product gases adversely affects the sulfur yield, a gas scrubber (raw water spray tower) is included to improve the yield by cooling the gases. The water content before water scrubbing is 60.2%. If adiabatic cooling were achieved, the following water contents would result:

Condensing Temperature (°F)	%H_2O (by volume)
100	6.5
120	11.5
140	19.7
160	32.2
180	51

8.5.11 After the reduction of the water content by cooling, the gas is reheated to normal Claus temperature (440-460°F) for conversion of the SO_2 and H_2S to elemental sulfur. Two units are provided with interstage cooling (sulfur condenser) for sulfur removal to effect a satisfactory overall yield. Entrained sulfur mist is removed by a coalescer. The exit gas is heated to 1000°F for combustion of the H_2S (oxidation of H_2S to the less toxic SO_2), before venting through the stack to the atmosphere.

8.5.12 For design purposes SVH's of 500 and 2000 are assumed for the Intermediate Converter and Claus Units respectively.

8.6 West Process (Texas Gulf Sulfur) - Gas No. 1

8.6.1 U.S. Patent 3,199,965 depicts 8 flow diagrams adaptable to this process, as illustrated by Claim No. 1: "The process of reducing sulfur dioxide to sulfur comprising reacting sulfur dioxide with a gaseous hydrocarbon in the presence of a catalyst selected from the group consisting of activated alumina, bauxite, calcium sulfide and quartz to produce elemental sulfur and a gaseous mixture comprising hydrogen sulfide, carbonyl sulfide, carbon disulfide and sulfur dioxide with a mole ratio of sulfur compounds other than sulfur dioxide of about 2:1 and during said reaction maintaining the temperature below 1830°F, cooling said gases to about 730°F, and contacting them with a catalyst to thereby react the carbonyl sulfide and carbon disulfide with sulfur dioxide to produce additional sulfur, thereafter cooling to about 125°C to condense the free sulfur and separating the same to leave hydrogen sulfide and sulfur dioxide in approximately a 2:1 mol ratio, thereafter heating the remaining gas in the presence of a catalyst to from 390 to 525°F to react the remaining hydrogen sulfide and sulfur dioxide and thereafter cooling to about 260°F to condense the sulfur and separating the same". As noted previously, the product composition cited is practically an equilibrium mixture. The only space velocity indicated was for a laboratory experiment using an SVH=95. Since the patent indicated a need for converting CS_2 and COS in an intermediate converter, it can be inferred that they conducted some of their work at sufficiently high feed rates that equilibrium was not attained and large amounts of by-product COS and CS_2 were formed. Specified herein is an SVH of 500 at 1830°F with an activated catalyst since others have proved that thereby equilibrium usually prevails. The patent also infers that activated alumina can be damaged by heating above 1830°F.

8.6.2 For the purpose of this design, the feed gases to the reactor are assumed preheated to 1380°F. At this temperature, the reaction proceeds at a reasonable rate. At lower temperatures, the reactor would be functioning both as a preheater and as a reactor. Although the latter condition is acceptable, the feeling is that for preliminary process visualization, one should keep the operation of the preheater and reactor as separate entities.

8.6.3 With the assumptions that the feed gas temperature is 1380°F and that the reaction takes place instantaneously at 1830°F, it is found that under adiabatic conditions the product gas temperature will reach 2080°F. It is further assumed that the maximum temperature of the catalyst must be kept below 1830°F to insure prolonged catalyst life. The use of a single high temperature reactor is not desirable since it appears difficult to provide internal cooling. Because higher feed ratios (greater than 2 to 1) decrease the adiabatic product gas temperature, values of outlet temperatures were computed for various SO_2/CH_4 ratios. See also plots of this data,Exhibits 8-24 through 8-31.

Product Gas Temperature		SO_2/CH_4 Feed Ratio *
(°C)	(°F)	
1136	2077	2.0
1108	2026	2.3
1083	1981	2.6
1052	1894	3.0
1015	1859	3.58

* This feed gas contains H_2O in the ratio of 1 part H_2O to 9 parts SO_2 (10% H_2O).

8.6.4 For process calculations, a feed ratio of 3.58 was selected which would result in an adiabatic gas temperature of 1860°F. Since the temperature profile was portrayed with and without a 7% heat loss (See Exhibit 6-37), it was decided to base the process calculations for the West Process (Case No. 2) assuming adiabatic conditions. This means that the maximum temperature to which the catalyst is exposed (except for local hot spots) should not exceed 1800°F. Computations for the Kulcsar Process (Case No. 4) however, are based on a 7% heat loss. The temperature differentials between adiabatic and 7% heat loss considerations are roughly equivalent to 110°F for the primary reactor and 200°F for the secondary reactor.

8.6.5 Although the proper feed ratio was established to give the desired reaction temperature, calculations revealed that if all the SO_2 were fed to the first reactor, the stipulated temperature (1830°F) would not be easily attainable in the second. For example, if the product gases resulting from feeding all the SO_2 to the first reactor were admixed with the balance of the methane (Temp.= 70°F), the temperature of the gas exiting the second reactor would be 1915°F. One of the easiest ways to decrease this temperature is to feed more cold feed gas to the reactor. Therefore, various proportions of the SO_2 (at the 3.58/1 ratio) were assumed fed to the first reactor and the resulting hot product gases mixed with the balance of the SO_2, H_2O and CH_4 (Temp.= 70°F) as feed to the second reactor. The temperatures of these feed gases and the resulting product temperatures were then determined:

Table XV

Percentage of SO_2 Fed to Primary Reactor	Temperature of Feed to Secondary Reactor		Product Gas Temperature from Secondary Reactor	
	°C	°F	°C	°F
100	880	1616	1046	1915
90	802	1476	995	1823
86	775	1427	975	1787
80	730	1346	944	1731

8.6.6 These results are plotted as Exhibit 8-5. The desired fraction of SO_2 and H_2O (Gas No. 1) to be fed to the first reactor should be approximately 90% in order to obtain the desired adiabatic secondary outlet temperature of 1820°F. These values were used for the Engineering cost estimate. The primary reaction temperature, the feed gas temperature to the secondary reactor, and the secondary reaction temperatures at 7% heat loss would be 1850°F and 1605°F respectively. It should be noted that this latter reaction temperature is the temperature level used for the Yushkevich Process and would be satisfactory at a lower throughput.

8.6.7 The temperature of the cooled product gases leaving the regenerators is shown as 950°F, assuming the sulfur species as S_2. Since S_2 polymerizes to S_6 and S_8 vapor upon cooling, the gas temperature should be higher because of the exothermic reactions:

$$3\,S_2 \longrightarrow S_6 \text{ and}$$

$$4\,S_2 \longrightarrow S_8$$

8.6.8 Trial-and-error calculations indicate that this gas should exit at approximately 1085°F with the sulfur composition at 10.86% S_2, 1.64% S_6 and 0.13% S_8 rather than the 15.71% S_2 which is used in the process projection. This type of detailed trial-and-error computation is deemed unnecessary for our present estimates but would be required for Phase III estimating.

8.6.9 For the Intermediate Converter, 100% consumption of the CO, H_2 and COS is envisioned as mentioned previously.

8.6.10 For the Claus Reactors, equilibrium conditions prevailing at a reaction temperature of 450°F are assumed. This equates to H_2S and SO_2 conversions of 75% and 25% in the first and second Claus units respectively. The data submitted to Engineering for the West Process Profile is shown in Exhibit 8-6. The lettered process streams, as listed on process profile, are shown on the associated flow sheet, Exhibit 8-1.

8.6.11 In retrospect, for the purpose of depicting the temperature profile with the primary reactor operating at about 1830°F after a 7% heat loss, it was necessary to feed all of the strong SO_2 to the first reactor at a 2.9 ratio. Under these heat loss conditions, the second reactor would be at 1255°F. The calculated temperatures for both reactors follow:

Table XVI

Process Stream	Condition	Temperatures °C	°F
Primary Reactor Exit	No Heat Loss	1061	1942
" " "	7% " "	996	1825
Secondary Reactor Feed	No Heat Loss	962	1764
" " "	7% " "	904	1659
Secondary Reactor Exit	No Heat Loss	1076	1969
" " "	7% " "	958	1756

8.7 Kulcsar Process - Gas No. 1

8.7.1 As indicated previously, Kulcsar presented the greatest amount of analytical data for the experimental work. These studies, using SO_2/CH_4 feed ratios of 2.6/1 and 3.1/1, clearly show the effect of varying both reaction temperature and retention time over a silica gel catalyst upon product composition. These experimental results demonstrate equilibrium performance except when the flow rates were excessive for the temperatures used and the reduction of SO_2 was incomplete.

8.7.2 Kulcsar's data also revealed the approximate maximum flow rates at varying temperatures at which equilibrium prevailed.

Temperature °C	Temperature °F	Maximum SVH (approximate)
1100	2012	6674
1000	1832	148-584
900	1652	148

8.7.3 It was further noted that the H_2S contents were lower than those at equilibrium and decreased with increasing SVH; that COS increases with increasing SVH and decreasing temperatures; that at 1110°F, COS is the only product formed.

8.7.4 Since the Kulcsar data encompassed only two SO_2/CH_4 feed ratios, namely 2.6/1 and 3.1/1, a natural question arises. "Will the same high conversions be obtained at a 2/1 feed ratio at 1100°C (2012°F) at this same extremely high space velocity (SVH = 6684)?" If so, a single reactor would be sufficient. For appraisal of this process using high strength SO_2 feed gas, it is assumed split feeds to a dual high temperature unit will be needed. A small reactor for the first unit at high SVH (6500) can be employed with a high (3/1) feed ratio. A larger unit is needed for the final reactor at an assumed SVH of 500. To complete this study, the temperatures of the feed gases at different feed ratios to reach an adiabatic product temperature of 1100°C (2012°F) were determined as follows.

Table XVII

SO_2/CH_4 Feed Ratio	Feed Temperature °C	Feed Temperature °F
2.0	765	1393
2.3	781	1448
2.6	801	1474
3.0	825	1517
3.58	856	1573

The equilibrium data required for these computations are shown graphically in Exhibits 8-24 through 8-28.

8.7.5 The equilibrium data required for analyzing Kulcsar's work (no water in the feed streams) are also included in Exhibits 8-29 through 8-31.

8.7.6 At the selected input (all of the strong gas and the CH_4 at a 3/1 ratio) to the first reactor, a problem arises in the second. Additional heat is required to produce 2010°F. On the Process Profile (Exhibit 8-11A) and the Flow Sheet (Exhibit 8-11), preheating the methane alone fulfills the heat requirements. In actual plant operation this would not be possible, since the methane would break down to carbon and deposit in the heat exchanger. Some process gas, steam, or possibly recycled stack gas could be employed as a diluent to minimize or prevent such a carbon deposition. Oxygen (air) could be injected into the reactor and heat of combustion would raise the temperature to the desired level.

8.7.7 Feeding 25% of the preheated strong gas (1650°F) to the second reactor would still require methane to be preheated to 1600°F. This slight benefit of lower methane temperature did not seem worthwhile.

8.7.8 In order to preheat the methane, a natural gas fired preheater is used. It is not feasible to use the hot product gases from the reactors as a heating medium. A conventional metal heat exchanger cannot be employed, since these gases are too corrosive for the metals generally considered for such service. Using two additional regenerators for this purpose would be economically impractical because of the increased fixed capital cost.

8.7.9 For the Claus units, the minimum SO_2 (after oxidation of the residual H_2S and S leaving the final sulfur condenser) that might be achieved is shown. Two water scrubbers were used, the first to cool the feed gases to the first Claus unit to 150°F, the second to cool the gases to the next unit to 130°F. The latter scrubber would drop the water content from 42% to 10%. 96% conversion occurs in the first Claus and 82% in the second. The SO_2 in the final plant effluent would still be about 1%, since the only diluents are by-product CO_2 and the small amount of H_2O remaining. The visualization of two Claus units is probably unrealistic. Because of the high H_2S and SO_2 content of the initial feed gas, the heat of reaction would raise the reaction temperature considerably above the stipulated 450°F. Accordingly, the conversion would be significantly lower. In practice, since the plant feed gas comes from an upstream collector system, one might try to use a single Claus unit, or two at most, and then vent the exit gases back to the stack of the upstream system.

8.8 ASARCO Process - Gas No. 1

8.8.1 The literature provided a good description of the construction and operation details of Asarco's semi-commercial unit (5 tons of sulfur daily capacity) which operated starting in 1940 for 4 1/2 years on copper smelter gases containing 5-8% SO_2 and 9-12% O_2. The reduction was accomplished on a firebrick bed at temperatures of at least (It was indicated that at lower temperatures carbon was produced, which contaminated the sulfur).

8.8.2 Since Fleming and Fitt (76) provided no analytical or rate data, equilibrium must be assumed in the high temperature reaction. A reasonable flow rate is selected since no data is available on non-catalyzed reactions at elevated temperatures. Refractory checkerwork may or may not act as a catalyst.

8.8.3 Averbukh's data confirmed the kinetics of the reduction at 2282°F by showing that equilibrium resulted at SVH's ranging from 58 to 960, but did not prevail at an SVH of 1870. (A considerable amount of CS_2 formed, accompanied by a substantial decrease of H_2S).

8.8.4 Averbukh also stated that "Separation of carbon in the reaction zone was observed at 900°C (1650°F), less frequently at 1100°C (2012°F), and was not observed at all at 1250°C (2280°F)".

8.8.5 However, Asarco indicated the need of an Intermediate Converter for conversion of the by-product COS. It must be assumed that Asarco operated the plant at rates in excess of those under which equilibrium would result.

8.8.6 This process is the only one which demonstrates the use of a single reactor. The high operating temperature, 2280°F (1250°C), of the Asarco Process produces larger proportions of carbon monoxide and hydrogen than the other processes which employ lower temperatures. This, in turn, results in a lower heat of reaction; hence, the feed gases must be preheated considerably more than for the other cases. For an adiabatic reaction temperature of 2280°F, the resulting temperature in keeping with a 7% heat loss would have been 2155°F. For a reaction temperature of 2280°F after a 7% heat loss, the adiabatic temperature would have been 2425°F. For process evaluation, a reaction temperature of 2280°F with the 7% heat loss was assumed.

8.8.7 This creates an interesting situation in that the preheating of the feed gases from ambient conditions (120°F) to the required 1990°F in the regenerators necessitates cooling the product gases from 2280°F to 347°C (660°F), assuming no polymerization of S_2 to S_6 and S_8. This temperature is less than the desired feed temperature (735°F) for the Intermediate Converter. If the reaction temperature had been 2155°F, as explained above, the temperature of the product gases leaving the regenerators would have been 480°F. The conversion of S_2 to S_6 and S_8 would have been barely sufficient to provide the additional heat necessary for the feed gas to the Intermediate Converter. The heat exchanger in the flow sheet (Exhibit 8-7) is described as a heater. In actual practice, it probably would be a gas cooler.

8.8.8 Claus conversions are the same as those shown for the West Process, namely 75% and 25%. The pertinent Process Profile is attached as Exhibit 8-7A. A spray tower is included, which should give higher Claus yields than those shown in the Process Profile.

8.8.9 In conclusion, no significant difference in the process profiles and flow sheets for the three catalytic methods appears, in spite of the significant differences in product compositions and heat balances. The Asarco Process has three inherent advantages over the catalytic processes:

- a single high temperature reactor suffices
- no catalyst is required
- the process has been adequately demonstrated in a semi-commercial unit.

8.8.10 The four process profiles on reduction of strong gas that we reviewed previously (Yushkevich, West, Kulcsar, and Asarco) entailed three major differences: a non-catalytic reduction (Asarco), catalytic at moderate space velocities (Yushkevich and West) and catalytic at high space velocities (Kulcsar). In view of these variances, both non-catalytic and catalytic processes for Gas 2A and Gas 2B were examined. Because of the obviously high costs entailed in reduction of Gas 2C or Gas 3 with methane, detailed studies were not justified. All process profiles include heat losses of 7% at each stage of the operation where temperatures exceed 1000°F.

8.9 No. 2A Smelter Gas

8.9.1 Case 3a - Asarco Process

Since it is predicted that this gas would discharge from the converters at 2300°F, particles of entrained gangue would be in the molten state as a mist. Normally, air is admitted to quickly cool the gas and solidify these particles as dust. At the temperature stipulated (2300°F) for Gas 2A (without air dilution), some of the molten particles would deposit and fuse on the walls of the flues, if some means of rapid cooling were not employed. It is proposed to cool the gas to 2000°F by injecting water into the gas stream near the converter exits (see Flow Sheet, Exh. 10-1, Process Profile, Exh. 10-2). If the fusion point of the entrained gangue is below 2000°F, cooling to lower gas temperatures would be required. The flues between the converter and the first waste heat boiler would be refractory lined.

8.9.2 The tubes in the two waste heat boilers can be fabricated from carbon steel. In the first unit, corrosion by SO_2 in an oxidizing atmosphere is negligible at the normal temperatures (600-800°F) employed in the water wall section of the boiler. In spite of the extreme corrosivity of the sulfur-bearing product gases (reducing atmosphere), carbon steel tubes may be used in the second boiler since the wetted wall temperatures are 545°F for saturated 1000 psia steam and a probable upper limit of about 800°F for superheated 1000 psia steam.

8.9.3 Since dust would probably impair the operation of the Intermediate Converter and/or Regenerator, its removal by a cyclone and an electrostatic precipitator is projected.

8.9.4 The high temperature reduction has been well demonstrated in Asarco's pilot plant, as reported in the literature. However, no details on reactor size, space velocities or product compositions were disclosed. In addition to the advantage of negligible catalyst cost (the reactor is filled with refractory checkerwork), the pressure drop is negligible. Therefore, this process shows less H.P. steam consumption (the process blower is driven by a steam turbine) than catalytic processes because of the lower pressure drop in the reactor.

8.9.5 It is assumed that the Intermediate Converter and Claus units in all methane reduction process profiles produce product gases at equilibrium compositions, so these units and the product gases from each unit will be quite similar. In earlier work, it was assumed that COS, H_2, and CO were completely consumed in the Intermediate Converter. At equilibrium, this is a reasonably correct assumption. In addition, the equilibrium studies for this case show that 11% of the H_2S and SO_2 is converted by the Claus Reaction to sulfur in this reactor. In the first Claus unit an 83% conversion at 450°F ensues. However, in the second unit, the effect of the high water content at low SO_2 and H_2S feed compositions is very pronounced; the conversion is only 35% at 425°F. At 450°F, the conversion would be 21%. A very low sulfur condensation temperature, 250°F is specified in order to minimize sulfur carry-over in the exit gases. The sulfur concentrations that are shown on the process profiles are conservative. Commercial Claus units with sulfur condensers running at 280-290°F can maintain the sulfur concentration at approximately 0.01% S_8 (0.04% as S_2). Sulfur mist is removed by the Coalescer.

8.10 Case 3b - West Process

The reaction for the West Process takes place at 1832°F. (See Process Profile, Exh. 10-3), which is substantially lower than the 2282°F temperature for the Asarco Process. At this temperature, the relative amounts of H_2 and CO are substantially lower and the heat of reaction is moderately higher than the corresponding values for the Asarco Process. The pressure drop through the West reactor is higher as noted previously. There appear to be no other significant differences in the two processes.

8.11 No. 2B Smelter Gas

8.11.1 Case 4a - Asarco Process

The simplest "Asarco Process" would be the adiabatic combustion of the oxygen and SO_2 with methane. The adiabatic reaction temperature is approximately 3010°F. With a 7% heat loss, the temperature would be about 2810°F. From a corrosion viewpoint, such temperatures may be border line for the use of any conventional metals in a waste heat boiler, so this approach was not used. Undoubtedly ceramic tubes could be used for the high temperature zone and conventional materials for lower temperatures.

8.11.2 The tentative plant design finally adopted was the same for all three processes (see Flow Sheet, Exh. 10-4). Since equilibrium conversions were assumed for the high temperature reactors, the intermediate converters and the Claus converters, the only differences that exist in the process profiles (see Exh. 10-5 Process Profile - Asarco Process; Exh. 10-6, Process Profile - West Process; Exh. 10-7, Process Profile Kulcsar Process) are the properties of the gases leaving the first waste heat boiler and the primary reactors. These differences are summarized in Table XVIII to illustrate the effect of increasing reactor temperatures.

Table XVIII

Item	Case 4b West	Case 4c Kulcsar	Case 4a Asarco	Trend of Variables
Gas Temp. (°F) from W. H. B. (1)	1854	2093	2457	Increases
Gas Vol. MACFM at T°F, from W. H. B. (1)	354	391	447	Increases
Gas Temp. (°F) from Reactor (2)	1832	2012	2282	Increases
Gas Vol., MACFM at T°F, from Reactor (2)	370	398	443	Increases
Reactor Product Composition (2) (Volume %)				
SO_2	1.26	1.53	2.07	Increases
H_2O	16.8	16.6	16.3	Decreases Slightly
H_2S	1.01	0.765	0.473	Decreases
CO_2	8.60	8.12	7.24	Decreases
CO	0.675	1.14	1.98	Increases
S_2	0.847	0.826	0.697	Decreases
H_2	0.798	1.14	1.68	Increases
COS	0.027	0.025	0.018	Amount Small Decreases
N_2	70.0	69.8	69.5	Practically Constant - Decreases because molar ratios of reaction products change.

(1) Process Point C.

(2) Process Point E.

8.12 Case 4b - West Process

8.12.1 The West Process employs a catalyst preferably activated alumina. Catalyst increases the rate of reaction so that lower temperatures (about 1800°F) can be employed, particularly desirable when the gases must be heated up by gas-fired preheaters. In this case, no apparent advantage in using a catalyst is seen, provided the reaction rates for the two processes are reasonably close to the assumed rates. Moreover, catalyst costs for the West Process might be excessive if catalyst life is short.

8.13 Case 4c - Kulcsar Process

8.13.1 This process was included in our studies since the reaction rates cited by Kulcsar using silica gel catalyst were extremely high and the use of a smaller reactor or reduced power costs might provide an economic advantage. Pressure drops and blower horsepower (See Exh. 10-8) were calculated for the Asarco, West, and Kulcsar Processes based on the visualized process designs. Two cases were computed for the Kulcsar Process to show the effect of the possible variation of pressure drop and reactor size as compared to the same parameters for the West Process. The first Kulcsar design, Kulcsar (1), Exh. 10-8, was used for the economic evaluation to note the effect of reactor size. No economic advantage was apparent. One might also predict that the life of silica gel catalyst would be less than that of activated alumina because of the higher reaction temperature.

Incomplete Information - Methane Reduction of SO_2

8.14 Reaction Kinetics

8.14.1 The design of a commercial plant based on present knowledge acquired from the literature, would be very risky because of the lack of important kinetic data. There is no reaction rate data on the Asarco Process. It appears that at the high temperatures, (2280°F and above) reaction might take place on the hot checkerwork surface at a satisfactory rate on a SVH of 500-1000. It is important to ascertain the amount of COS formed since an Intermediate Converter might not be required if equilibrium (negligible COS) were attained. In addition, if one considered the alternate method of using silica gel in a single reactor, one does not know the reaction rate at a 2/1 feed ratio of SO_2 to CH_4.

8.15 Materials of Construction

8.15.1 The hot product gases will be extremely corrosive. It is important to know whether any metals are resistant to these severe conditions. Presently, it must be assumed that the valves in this high temperature service can be water-cooled and/or refractory-lined to satisfactorily minimize corrosion hazards. Operating maintenance must be moderate in this type of plant if estimated costs are to apply.

8.16 Intermediate Converter

8.16.1 Complete reaction of hydrogen, carbon monoxide and carbonyl sulfide (COS) is assumed at the prevailing conditions. This appears optimistic since complete conversion of all three does not seem likely. Since incomplete conversion means incomplete utilization of the full reducing power of the methane, increased methane costs would result.

For example, a 10% loss in reducing power is equivalent to an increased operating cost of $0.38 per ton of sulfur. For detailed cost estimates, it is important that the yields and reaction rates for the Intermediate Converter be known.

8.17 Catalyst Life

18.17.1 For the Yushkevich, West and Kulcsar Processes, the life of the bauxite, activated alumina or silica gel is assumed to be at least six months. It is well known that these materials shrink severely and exhibit a very substantial reduction in surface area at 1830°F to 2280°F. If such physical changes decrease catalyst life significantly, a commercial facility using a catalytic reactor system might not be economically feasible. In that case, one would have to use the Asarco Process.

9.0 REFORMER GAS REDUCTION

INTRODUCTION

9.1 This study was made to provide a comparison of the direct reduction by methane and the reduction with reformed methane.

The extra step of reforming the methane must be justified by factors such as the following:

Process simplicity
Low reduction temperature
Long catalyst life
Less demanding on materials of construction

9.2 The reduction with reformed methane represents something near the ultimate in process simplicity. Refer to Exhibits 10-9 and 10-10. The reformer gas consisting of a mixture of H_2 and CO as the reductant is mixed with the SO_2 gas stream and passed over a catalyst bed.

9.3 The temperature is moderate and is a compromise between reaction rate and equilibrium degree of completion. The rate is favored by higher temperature and the equilibirum is favored by lower temperature. At 930-1110°F the H_2 and CO can be reacted virtually to completion.

9.4 The H_2, CO reaction with SO_2 is considerably more exothermic than the reaction with CH_4. The high heat release requires two converters with interstage cooling in order to obtain a favorable temperature leaving the "B" converter catalyst bed.

9.5 The gas leaving the "B" converter is cooled and passed to a Claus unit for conversion of residual H_2S and SO_2 and the recovery of sulfur as a product.

9.6 NO. 2A SMELTER GAS

Two cases, 5a and 5b, are included covering conditions as follows.

9.6.1 Case 5a is based on mixing hot reformer gas and the smelter gas then cooling to 572°F before entering the first converter. The gas leaving the "A" converter is cooled and passed thru "B" converter and cooler. Sulfur is recovered from the "B" cooler and the gas is passed to a Claus unit for final sulfur recovery. Refer to Exhibits 10-9, 10-10, 10-11, and 10-12.

9.6.2 In case 5b, a large part of the water in the reformer gas is removed by cooling the gas under the 10 atmospheres reformer pressure. The low water reformer gas is then mixed with the smelter gas and processed the same as indicated for Case 5a. Refer to Exhibits 10-11 and 10-12.

9.6.3 The purpose of the water removal was to define the yield improvement due to a better recovery in the converters and in the Claus unit on the lower water content gas. The water removal is a very simple step and would probably be justified if reformed methane were used in a sulfur recovery process. The capital cost of the plant is based on equipment and sizing as follows.

9.6.4 The reformer unit is based on an estimate furnished by the Selas Corporation of the cost of a package unit specified as follows:

(a) outlet pressure 165 PSIA
(b) reformer temp. 1650°F
(c) Water/CH_4 ratio 3.0:1
(d) inlet sulfur 1/2 ppm max.
(e) available steam pressure 250 P.S.I.G.
(f) 2000 SCFM CH_4 feed.

Reformer conditions were selected to limit the CH_4 at equilibrium to 2.0%.

9.6.5 The "A" & "B" Converters are based on 3.0 ft./sec. superficial velocity and 2.0 sec. superficial residence time. SO_2 conversion leaving "B" Converter is estimated as follows:

Case 5a	23.7%
Case 5b	40.0%

9.6.6 The Claus Units consist of two beds with interstage sulfur removal. Operating conditions are defined as shown in Table XIX below.

TABLE XIX

Case	5a	5b
1st stage		
Temp. in °F	400	400
Temp. out °F	540	520
Conversion %	77.0	80.0
2nd stage		
Temp. in °F	400	400
Temp. out °F	429	436
Conversion %	65.3	66.0
Overall System Conversion	93.91	95.88

9.7 Conclusions

9.7.1 The reduction system based on reformed methane (or other hydrocarbons) has several attractive features which may make additional Phase I type studies on other gas streams desirable.

9.7.2 Cost studies show that the cost of the reformer unit can be justified by counterbalancing cost savings.

9.7.3 The minimum catalyst temperature is in the 1000-1100°F temp. range while temperatures above 1800°F are required in the direct reduction with methane.

9.7.4 The catalyst temperature is far below the maximum allowable for activated alumina catalysts and the catalyst should have a very long life.

9.7.5 At SO_2 levels much under 10%, the direct reduction by methane would not be thermally self-supporting and would require methane combustion to maintain the required conditions. This would not be the case with reformed methane where the reaction proceeds at a comparatively low temperature and with a strong exothermic effect. With reformed methane, it is likely that a process could be devised to work on only 2 or 3% SO_2 in an inert gas stream.

9.7.6 By inference, the same advantages noted for reformed methane as reductant for gas 2A should apply as well to the gases 1, 2B and 2C.

10.0 REDUCTION WITH H_2S

INTRODUCTION

10.1 The method of sulfur recovery by the reaction of H_2S and SO_2 is an effective system. There are several process modifications of the "Claus" Process but all are based on the following general reaction:

$$2H_2S + SO_2 \longrightarrow \frac{3}{n} S_n + 2H_2O$$

10.2 The common application of the Claus process is in the recovery of sulfur from H_2S gas streams. One third of the H_2S is burned with air to give H_2S and SO_2 in the required ratio.

10.3 The gas is passed over an activated alumina or bauxite catalyst and sulfur is recovered from the converted gas. Recovery efficiency is decreased by dilution with inert gas and by increasing H_2O content.

10.4 The recovery of sulfur from an SO_2 gas stream requires the generation of the required H_2S from a recycle of a part of the recovered sulfur. In theory two thirds of the sulfur is recycled but in a practical system the recycle ratio is significantly higher than 2/3.

10.5 Any H_2S reduction system will contain prohibitive quantities of H_2S in the Claus exit and will require combustion of this H_2S to SO_2 with supplemental fuel.

10.6 In the studies of H_2S reduction two fundamentally different systems are considered.

10.6.1 Claus vapor phase catalysis of $H_2S + SO_2$ in the nominal temperature range of 400-600°F where the converted sulfur exits as sulfur vapor. Applications herein relate to feed conditions not normally used in Claus plants.

10.6.2 Low Temperature Claus Catalysis of $H_2S + SO_2$ in a temperature range of about 150-350°F where the sulfur is retained on the catalyst as solid or liquid sulfur. This sulfur is removed and recovered in a separate regeneration step.

10.7 H_2S Generation Three methods for generating the required H_2S have been considered. Two of these produce H_2S from recycled sulfur according to overall reactions as follows.

10.7.1 Reformed hydrocarbons and sulfur vapor

$$2H_2 + S_2 \longrightarrow 2H_2S$$

$$2CO + S_2 + 2H_2O \longrightarrow 2H_2S + 2CO_2$$

10.7.2 Methane + sulfur vapor + water

$$CH_4 + 2S_2 + 2H_2O \longrightarrow 4H_2S + CO_2$$

10.7.3 The third system is based on the reaction of methane with SO_2 and might be applicable to a strong SO_2 gas.

$$4SO_2 + 3CH_4 \longrightarrow 4H_2S + 3CO_2 + 2H_2O$$

10.8 Generation of H_2S from CH_4 and SO_2 The first step was to develop the equilibrium information on the SO_2-CH_4 system covering a range in the SO_2/CH_4 ratio at R = 0.8, 0.9, 1.0, 1.1, and 1.2. This equilibrium data is presented in Exhibits 6-1 thru 6-5.

10.8.1 The H_2S reduction is based on the premise that the H_2S generation step must be justified by a balancing process simplification. The simplification in this case is the ability to mix the H_2S and a low oxygen SO_2 stream and directly enter a Claus unit. The basis seems justified in view of the demonstrated ability to reduce SO_2 with either CH_4 or hot carbon in a direct process. If the simple process visualized does not result, and if additional units, such as intermediate catalytic beds are required, then the interposition of an H_2S synthesis step is redundant. This means that to the extent that CO and H_2 are formed, the CH_4 efficiency will suffer. Conflicting information on the extent to which COS reacts in a conventional Claus unit was found and its real extent should be resolved in a Phase II study.

10.8.2 Inspection of Exhibits 6-1 thru 6-5 indicate a substantial formation of H_2 and CO over the entire temperature range. In order to define the CH_4 utilization, an Efficiency Factor "E" was defined which is based on the stoichiometry of the process and credits the methane only with the formation of sulfur or H_2S. This factor will be unity when there is no formation of CO, H_2 or COS. See Exhibit 6-6. It is clear that in the range of ratios around 1.0 where H_2S synthesis would be conducted that the CH_4 efficiency is low. The conversion of SO_2 to H_2S is good but CH_4 is wasted in the production of CO, H_2 and COS.

10.8.3 The curve R = 2.0, Exhibit 6-6, represents the reaction

$$CH_4 + 2SO_2 \longrightarrow S_2 + CO_2 + 2H_2O$$

This is the condition near optimum for the direct reduction of SO_2 by CH_4. At high temperatures there is a penalty due to the formation of H_2 and CO which may justify an intermediate catalyst bed to improve the CH_4 efficiency but operation in the H_2S generation area around R = 1.0 would impose a major duty on the intermediate reactor with no process justification.

10.8.4 Our conclusion from this study is that the generation of H_2S from strong SO_2 and CH_4 is performed at a low CH_4 efficiency, and that the direct reduction of SO_2 with CH_4 at about R=2.0 is the preferred route over any H_2S reduction process.

10.9 Generation of H_2S from CH_4 and S_2 Study of the system based on the second reaction $CH_4 + 2S_2 + 2H_2O \longrightarrow 4H_2S + CO_2$ was next undertaken. The first step was to obtain comprehensive equilibrium data over the temperature and reactant ratio range of interest. The temperature covered the range 600-2400°F in 100° increments. The reactant ratios included covered ten combinations as follows.

TABLE XX

Exhibit	6-7	6-8	6-9	6-10	6-11	6-12	6-13	6-14	6-15	6-16
R H_2O/CH_4	2.0	2.0	2.2	2.2	2.2	2.2	2.2	2.2	.24	.24
R S_2/CH_4	2.0	2.2	2.0	2.2	2.4	2.7	3.0	3.3	2.0	2.2

The results of these equilibrium determinations are presented in Exhibit 6-7 & 6-16 attached.

10.10 In the same manner as followed in the previous H_2S synthesis from SO_2 and CH_4, a methane efficiency factor for the synthesis from CH_4-S_2-H_2O was calculated. This factor credits the CH_4 only for the H_2S produced and will be 1.00 when no CO, H_2, COS or CS_2 is formed. These CH_4 efficiency factors are presented in Exhibits 6-17, 6-18, and 6-19 attached.

10.11 At a specific temperature it is possible to present all the information concerning the effect of reactant ratios on a single chart. Later in this report, Case 6-1 based on such a synthesis at a reaction temperature of 1300°F is presented. Exhibit 6-20 shows the effect of reactant ratios on the CH_4 efficiency at equilibrium at 1300°F.

10.12 Reduction of SO_2 with H_2S - Normal Claus

In order to gain a perspective on the merit of this system of generating H_2S, an overall sulfur recovery system using this method was projected. This Case 6-1 is useful as an initial outlook on the system, but it should not be considered optimum. It should give a feel for the important advantages and deficiencies in the scheme.

10.13 Following are the major premises on which the visualized process is based:

A 1300°F H_2S synthesis temperature
Stoichiometric reactant ratios
Recycle sulfur revaporized by direct CH_4 - air combustion gas
Rating based on 20,000 SCFM converter gas 2A
10.0 sec. superficial residence time in the H_2S converter
4% heat loss in sulfur vaporizer
7% heat loss in H_2S converter
Complete CH_4 reaction

A diagram of the plant layout is given in Exhibit 6-26. The details of the projected process profile as letter keyed to Exhibit 6-26 is detailed in Exhibit 6-21-A.

10.14 Comments based on this preliminary study are as follows

10.14.1 The H_2S synthesis system appears practical. The methane consumption should allow for the CH_4 efficiency as well as for unreacted CH_4.

10.14.2 Work on reactant excess shows that an excess of H_2O improves the efficiency. This improvement would have to be balanced against the adverse effect of H_2O on the Claus unit recovery; however, such a study was deferred.

10.14.3 Excess S_x also improves CH_4 efficiency and in this case one can make definite judgment. Referring to Exhibit 6-20, note that increasing the S_x ratio to a 10% excess while holding the H_2O ratio at theory increases the CH_4 efficiency from 88.90% to 90.35%. By Exhibit 6-21-A, the CH_4 feed to the H_2S synthesis is 1870 SCFM. At the higher efficiency the CH_4 flow is 1840 SCFM or a reduction of 30 SCFM. The CH_4 required to vaporize the recycle sulfur is 880 SCFM. The 10% excess S_x would require a 10% increase in the CH_4 to the vaporizer equal to 88 SCFM. The net effect would be an increase in CH_4 use equal to 88-30 or 53 SCFM. The use of excess sulfur is unjustified due to the energy requirement of the revaporization.

10.15 The low sulfur yield of 82% points up an inherent weakness of a process based on a heavy sulfur recycle when the system is backed up by a conventional Claus unit. In the subject case, the Claus unit yield at equilibrium is slightly over 98%. In this study 96% was used, a yield in line with the performance of commercial units. The low yield is due to the fact that the Claus unit has to recover almost 4 parts of sulfur for each part removed as product. This multiplies the loss in direct ratio to the recycle factor. It appears that a conventional Claus unit is not efficient enough to back up a system where a heavy sulfur recycle is involved. Probably an operating efficiency of at least 99% in the sulfur recovery unit would be required to hold the exit sulfur to a satisfactory value.

10.16 In this study it is assumed that the COS is reactive to the same degree as H_2S in the Claus unit. This is a point on which authorities differ. If the COS does not react, the sulfur yield would decrease from 82% down to 73%.

10.17 Gamson and Elkins (CEP 49 No. 4 p. 11) show results on a 5.5% COS gas with 2.75% SO_2 giving around 90% reaction at a space velocity of 200 SVH.

10.18 Lepsoe (I & EC 32, No. 7 p. 916-917) gives some information on the COS-SO_2 reaction at SVH about 60, providing much longer contact times than the SVH of about 2000, which is more normal in a Claus unit. Experimental work is required to determine the efficiency of COS utilization when reacting along with a larger amount of H_2S. The gas condition in Case 6-1 of 37.6% H_2S and 1.94% COS is a case in point. Equal COS reactivity is assumed, but there is really no good information to justify this assumption. The large influence on the sulfur recovery efficiency makes it necessary to study this system.

Work on Case 6-1 leads to the tentative conclusion that a system based on the generation of H_2S by recycled sulfur product can be justified only where there is no alternate. With low oxygen gas and about 10% SO_2 or more, the direct reduction with CH_4 or reformed CH_4 will be preferable.

11.0 LOW TEMPERATURE CLAUS PROCESS

INTRODUCTION

11.1 The Low Temperature Claus process is distinguished from the conventional Claus by operation at a temperature low enough so that oxygen does not react. The temperature limit is considered to be about 390°F and the bulk of recent work has been done at about 260°F

11.2 The sulfur produced is retained on the catalyst as solid or liquid sulfur. Information available at this time suggests that the catalyst may be loaded with about 20% sulfur by weight before its activity as a catalyst is significantly impaired. Such a sulfur loading is here assumed, but it must be confirmed that this loading is attainable in a practical operating unit.

11.3 The sulfur is recovered and the catalyst regenerated by heating and purging with an inert or reducing gas. Recent work indicates that catalytic activity of activated alumina catalysts is lost if the inert purge does not contain a small content of reducing gas such as H_2S.

11.4 Studies on Catalysis and Adsorption

11.4.1 Preliminary calculations indicated to us that a serious sulfur loss could occur if the sulfur on the catalyst was exhibiting its normal sulfur vapor pressure in the temperature range of the low temperature Claus Process. This means that if a large loss as a sulfur vapor is to be avoided, the catalyst must act as a sulfur vapor absorbent as well as a catalyst. The proportional effect of sulfur vapor pressure is greatest on weak gas such as power house stack gas and becomes a much less important factor on smelter gas and other stronger SO_2 gas streams.

11.4.2 As an initial step in the Low Temperature Claus studies, calculations designed to separate the catalytic and the adsorptive function of the catalyst were made. These studies are based on a calculation of the conversion and process recovery under the limiting conditions of 100% sulfur vapor adsorption and zero adsorption. An actual process will operate somewhere between these limits. The results of these calculations are summarized in Tables 7-1, 7-2, 7-3 and Exh. 7-1 and 7-2 which are included at the end of this section.

11.4.3 The equilibrium calculations are unusual in that where equilibrium between sulfur and its vapor is assumed, the S_8 fraction in the gas is determined by sulfur vapor pressure and not by material balance. S_6 will be present to a small extent in the higher temperature range considered here, but the vapor is assumed to be entirely S_8. The method used for these equilibrium calculations is detailed in section 13.0 of this report.

11.4.4 The first calculations (See Table 7-1 for summary of the results) cover the effect of conversion temperature on sulfur recovery. The adsorption effect was defined by calculating the sulfur recovery at both 100% adsorption and at zero adsorption where the sulfur on the catalyst would exert its full vapor pressure at the conversion temperature. The effect of the yield in the H_2S synthesis from recycled sulfur was evaluated at 100, 95, and 90%.

11.4.5 The H_2S synthesis efficiency yield factor brought into this study applies only to the sulfur balance. In any practical H_2S synthesis from sulfur, with the single exception of a pure hydrogen reactant, the gas will contain COS and CS_2 to some extent. The H_2S efficiency would be 100% where the COS and CS_2 react with SO_2 to form sulfur to the same extent that H_2S reacts. Formation of CO and H_2 increase the reductant consumption but are not considerations in this study.

11.4.6 Kinetic factors are involved here and the question which must be answered is whether or not the COS and CS_2 react fully in the residence time provided on the catalyst. If these reactions with SO_2 are slower than the H_2S reaction, then either a larger catalyst charge must be provided in the unit or a reduction in H_2S efficiency be accepted and either of these alternates would involve an economic penalty.

11.4.7 The literature carries references which indicate that in a conventional Claus unit, COS may react at a substantially lower rate than H_2S. The conditions in the low temperature Claus are radically different and the kinetic comparison may be quite different. The process yield depends on the reactivity of the COS and CS_2 and the kinetic factors require explicit study. The term "recoverable sulfur" used in Table 7-1 is the sulfur entering as H_2S and SO_2 trapped on the catalyst as solid or liquid sulfur above the amount of recycle sulfur required to generate the H_2S. This "recoverable sulfur" is expressed as a percent of the SO_2 feed. In order to show the trend of the effect of the variables, data for the performance at a 95% synthesis yield is plotted in Exhibit 7-1.

11.4.8 The next study was directed to the effect of the water content of the gas on sulfur recovery. At the higher temperatures of a conventional Claus unit, the recovery is strongly dependent on water content. At the low temperature under consideration here, the very high values of the equilibrium constant should greatly reduce the adverse effect of water. The results of these calculations covering the range from no water up to a pure steam atmosphere, is shown in Table 7-2. These results are plotted and shown as Exhibit 7-2. Curves B, C, D, and E show % Recoverable Sulfur vs. % H_2O Content. As might be expected, the conversion is not radically decreased by even large water ratios.

11.4.9 Two final calculations were made as follows. Both the conversion loss and loss due to sulfur vapor pressure will increase as the SO_2 content of the incoming gas is reduced. For all calculations up to this point, gas No. 3 was used, carrying the following analysis:

0.30	SO_2
7.25	H_2O
14.70	CO_2

2.80	O_2
0.05	NO_x
74.90	N_2

For these calculations, the effective analysis used was:

0.30	SO_2
7.25	H_2O
92.45	Inert

11.4.10 Also calculated was a composition with 0.20% SO_2 instead of 0.30%. The results are in column A of Table 7-3. These results may be compared with corresponding gas at 0.30 SO_2 by referring to the 7.25% H_2O column in Table 7-2. As would be expected, the conversion and sulfur recovery are substantially lower at the lower SO_2 level in the stack gas.

TABLE 7-1

LOW TEMPERATURE CLAUS

EFFECT OF TEMPERATURE AND CATALYST ADSORPTION

Temp. °C	100	120	140	160	170	180
Temp. °F	215	248	284	320	338	356
Conversion %	99.00	97.51	95.67	91.33	88.66	86.66
No Catalyst Adsorption						
% S as vapor	0.89	3.73	12.88	38.22	64.00	86.66
% S as liquid or solid	98.11	93.78	82.79	53.11	24.66	0.00
Recoverable Sulfur - no adsorption						
100% H_2S Synthesis Eff.	94.3	81.3	48.3	0.0	0.0	0.0
95% " " "	83.7	70.7	37.7	0.0	0.0	0.0
90% " " "	72.3	59.3	26.3	0.0	0.0	0.0
Recoverable Sulfur - 100% adsorption						
100% H_2S Synthesis Eff.	97.0	92.5	87.0	74.0	66.0	60.0
95% " " "	86.3	81.8	76.3	63.3	55.5	49.3
90% " " "	75.0	69.8	65.0	52.0	44.4	38.0

Feed

SO_2	0.30%
H_2S	0.60%
H_2O	7.25%
N_2	91.85%

TABLE 7-2

LOW TEMPERATURE CLAUS

EFFECT OF WATER

120°C (248°F)

% H_2O	0	7.25	15.0	99.1
Conversion %	99.80	97.51	96.16	86.66
No Catalyst Adsorption				
% S as vapor	3.73	3.73	3.73	3.73
% S as liquid or solid	96.27	93.78	92.43	82.93
Recoverable Sulfur - no adsorption				
100% H_2S Synthesis Eff.	88.8	81.3	77.3	48.8
95% " " "	78.1	70.7	66.6	38.1
90% " " "	66.8	59.3	56.5	26.8
Recoverable Sulfur - 100% adsorption				
100% H_2S Synthesis Eff.	99.4	92.5	88.5	60.0
95% " " "	88.7	81.8	77.8	49.3
90% " " "	77.4	69.8	66.5	38.0

Feed

SO_2	0.30
H_2S	0.60
N_2+H_2O	99.10

TABLE 7-3

LOW TEMPERATURE CLAUS

CONVERSIONS AND RECOVERABLE SULFUR

120°C (248°F)

	A	B
Feed % SO_2	0.20	0.30
% H_2S	0.40	0.45
% H_2O	7.25	7.25
Conversion %	96.35	89.20
No Catalyst Adsorption		
% S as vapor	5.60	4.48
% S as liquid or solid	90.75	84.72
Recoverable Sulfur - no adsorption		
100% H_2S Synthesis Eff.	72.25	61.8
95% " " "	61.13	53.8
90% " " "	49.75	45.1
Recoverable Sulfur - 100% adsorption		
100% H_2S Synthesis Eff.	89.05	73.0
95% " " "	78.05	65.0
90% " " "	66.55	56.3

A - Lower SO_2 in stack gas

B - H_2S at 75% of theory to reduce H_2S content in exit

11.4.10 The final set of calculations was made to define the degree to which the H_2S content in the vent gas could be reduced by running at less than theory on H_2S.

The feed gas is taken at 3000 ppm SO_2 with four ratios of H_2S/SO_2 as indicated. The H_2S and SO_2 content in the exit gas when operating over range of from 95.0 to 100% of equilibrium conversion is shown as ppm in the following tabulation.

TABLE XXV

Ratio H_2S/SO_2	2.0		1.5		1.0		0.5	
Percent of Equilibrium Conversion	H_2S	SO_2	H_2S	SO_2	H_2S	SO_2	H_2S	SO_2
100.0	130	65	37	769	26	1513	21	2260
99.9	136	72	41	771	29	1515	22	2261
99.5	160	80	59	780	41	1521	30	2265
99.0	189	90	82	791	56	1528	35	2268
98.0	248	124	126	813	86	1543	50	2275
97.0	306	153	171	836	115	1558	65	2283
96.0	365	183	216	858	145	1573	80	2290
95.0	424	212	262	881	175	1588	95	2298

11.4.11 The preceding tabulation indicates that the H_2S content in the stack gas exit cannot be reduced to an acceptably low value by operating at a reduced H_2S/SO_2 ratio even if equilibrium is attained. At lower conversions more typical of an actual operation the performance degrades rapidly.

11.4.12 Operation at ratios lower than theory result in a serious reduction of the recovery of SO_2. This means that while the H_2S may be reduced by such low ratio operation the SO_2 abatement becomes very poor.

11.4.13 This effect is shown by calculating the theoretical % sulfur yield obtained at different H_2S/SO_2 ratios and at different conversions. The following table is based on a 3000 ppm. SO_2 gas and tabulates the sulfur yield under different conditions. 100% adsorption of sulfur vapor is assumed.

% of Eq. Conversion	H_2S/SO_2 RATIO 2.0	1.5	1.0	0.5
100	94	73	49	24
99	83	71	46	23
95	79	62	41	21

11.4.14 From these preliminary studies on the Low Temperature Claus System as applied to stack gas, it is concluded that:

- The sulfur recovery is critically dependent on sulfur vapor adsorption by the catalyst.
- Operation at H_2S/SO_2 ratios less than theory will not result in an acceptably low H_2S content in the exit.
- The SO_2 pollution abatement will be adversely affected by H_2S/SO_2 ratios less than the theoretical 2.0.

11.5 LOW TEMPERATURE CLAUS PROCESS STUDIES

INTRODUCTION

11.5.1 This report will cover studies on this system as applied to the 2B and 2C Smelter Gas and to the #3 Stack Gas. Included are separate studies on the generation of H_2S gas from the CH_4-S_2 - H_2O reaction and from reformed CH_4. Two different systems for the regeneration of the sulfur loaded catalyst are included. The regeneration system to be used must remain open until the detailed processing requirements are defined for this operation.

11.5.2 It is recognized that the details of commercial unit visualizations, as called for in this contract must remain tentative until the process is fixed. There is a strong feeling, however, that such studies are essential in order to define the nature and magnitude of the full size unit scale up problems. Unless this is done, the pilot plant may not supply an adequate background for the engineering of a large unit.

11.5.3 Probably the best pilot plant design results from a scale-down from a visualized operable commercial unit rather than by a scale-up from bench work.

11.5.4 Studies herein are referred to a superficial contact time based on flow volumes expressed in SCFM at 1 atm. and 0°C. This space velocity is equivalent to a superficial contact time of 1.35 sec. Princeton Chemical Research (PCR) is now reporting a contact time based on the void volume of the catalyst. With the catalyst used in their run 64A, PCR reported a contact time of 0.76 sec.

11.5.5 In this study there are eight cases, with 4 cases being alternates for the first case, under the general subject of Low Temperature Claus, as follows:

Case 1 - 2B Smelter Gas - 70,000 SCFM

- H_2S Generation from Methane, Sulfur and Water
- 3 Stage Converter with External Intercoolers
- Tubular Sulfur-Heated Regenerator

Case 1-a - Same as Case 1, except moving bed regenerator, directly heated by inert gas

Case 1-b - Same as Case 1, except H_2S generation from reformed methane and sulfur

Case 1-c - Same as Case 1-b, except low temperature operation with cooling by water spray injection

Case 1-d - Same as Case -1b, except single stage adiabatic reactor based on vent gas recycle

Case 2R - Low temperature Claus applied to No. 3 stack gas

Case 2S - Low temperature Claus applied to No. 3 stack gas using switch reactors.

Case 6 Low temperature Claus applied to 2-c smelter gas

11.6 Case 1 - 2B Smelter Gas

- H_2S Generation from Methane, Sulfur and Water
- 3 Stage Converter with External Intercoolers
- Tubular Sulfur Heated Regenerator

11.6.1 An equipment layout diagram for Case I is shown as Exhibit 9-1 and the corresponding process profile, letter keyed to Exh. 9-1, is shown on Exh. 9-2. The H_2S is generated by the CH_4 - S_2 - H_2O reaction, $CH_4 + 2S_2 + 2\ H_2O \longrightarrow 4H_2S + CO_2$ and is identical with the same system shown in Exh. 6-26, except that the quantities are 92% of those shown in Exh. 6-21A.

11.6.2 The gas cooler is required to reduce the H_2S gas temperature to 257°F prior to mixing the smelter gas so that oxygen will not react.

11.6.3 The Claus converter is shown as a three stage moving bed unit with intercoolers. This is necessary since the adiabatic temperature rise in a single stage would result in a rise from 257°F to 531°F above which oxygen is stated to react. If oxygen started to react, a temperature excursion above 2300°F is possible, which could destroy the unit.

11.6.4 In the three stage unit, the maximum temperature is 372°F and the final equilibrium is established at 302°F Control is maintained by limiting the H_2S feed to each stage so that a superficial velocity of 4.0 ft./sec.

11.6.5 The assumption is made that the proposed moving bed system is operable. This would require demonstration in a pilot unit designed to study the material transport problem as well as the sulfur capture efficiency. A somewhat more detailed representation of the unit is shown in Exh. 9-3.

11.6.6 All calculations are based on a loading of 20% sulfur in the loaded catalyst. It is assumed that COS and CS_2 react to the same degree as H_2S.

11.6.7 The pressure drop on the converter is estimated at 177" H_2O and represents 5800 horsepower at an overall efficiency of 60% on the blower. Pressure drop estimates are based on a 4/8 mesh granular catalyst.

11.6.8 The use of three stages in the converter is arbitrary. The three stage concept results in a distribution of the adiabatic temperature rise of 274°F to give a maximum temperature of 372°F. If this maximum temperature proves to be higher than desirable then more stages would be required. A 5-stage converter operating on 2B smelter gas would allow a maximum temperature of 311°F. The staging will also depend on the SO_2 concentration in the smelter gas. A higher SO_2 content would require staging appropriate to the higher adiabatic rise in temperature.

11.6.9 The regenerator is detailed in Exhibit 9-4. The duty in the three stage Claus reactor is so different from that in the regenerator (chemical reaction and sulfur absorption vs. sulfur desorption) that a quite different design is indicated. No system based on switching from conversion to regeneration in the same unit seemed practical.

11.6.10 This regenerator is indirectly heated by sulfur vapor while the loaded catalyst moves down in tubes against an inert or reducing gas flow. The unit consists of 380 6" tubes each 27 ft. long. This unit gives a 1/2 hr. residence time. The superficial gas velocity is 0.365 ft./sec. and the heat flux across the tubes is calculated as 10.8 BTU/sq. ft./hr./°F.

11.6.11 The calculated pressure drop in this unit is 17.2" H_2O. The 514 SCFM purge gas recycle calculates to a power demand of only 3.3 horsepower at 60% efficiency on the recycle blower. This regeneration system will be attractive if a residence time of the order of 1/2 hr. proves satisfactory. If the residence time is several hours, this tubular type unit would be impractically large.

11.7 Case 1-a - Same as Case 1, except Moving Bed Regenerator Directly Heated by Inert Gas

11.7.1 An alternate regenerator design is shown in Exh. 9-5. This is based on heat supplied directly by heated inert gas. This unit is 20 ft. dia. and provides a 3.0 hr. residence in a 37 ft. deep bed. A superficial velocity of 4.0 ft./sec. is taken.

11.7.2 At 20% sulfur loading, this regenerator system appears operable. About 70% of the heat demand is taken by heating the alumina catalyst and at a sulfur loading significantly under 20%, the increased heat demand might make this type unit impractical. Such a decrease in sulfur loading would require an increase in the diameter of the unit and would lead to a decrease in the sulfur content in the recirculating gas leaving the unit below the already rather low value.

11.7.3 The pressure drop in this system using a 4/8 mesh catalyst is a prohibitive 900" H_2O. This design would be more practical if a larger size catalyst were used to reduce the pressure drop.

11.7.4 The process profile around this unit is as follows and refers to the letter coding on Exh. 9-5.

Table XXI

Process Point	Q_1	R_1	R_2
Temp. °F	500	932	257
Temp. °C	260	500	125
Vol. SCFM	76760	75000	75000
Vol. CFM at T°C	140900	213000	109400
% S_6	0.50		
% S_8	1.20		
% N_2	98.30	100.00	100.00

11.7.5 This design can provide an extended residence time if this factor is critical. The regeneration parameters will have to be defined precisely before a decision on regenerator design approach can be made.

11.8 Case 1-b - Same as Case 1, except H_2S Generation from Reformed Methane and Sulfur

11.8.1 Case 1-b in the low temperature Claus series covers the generation of the required H_2S from reformed methane:

$$CH_4 + H_2O \longrightarrow CO + 3H_2$$

The process profile for this system is shown as Exh. 9-6 and the corresponding equipment layout diagram is shown as Exh. 9-7.

11.8.2 The reformer gas composition shown represents reformer operation at the following conditions:

H_2O/CH_4 ratio - 3.0:1
Temp. - 1650°F
Pressure - 10 at.

11.8.3 The H_2O/CH_4 ratio selected above is required to limit the CH_4 at equilibrium to 2.0%. At lower water ratios, the equilibrium CH_4 increases rapidly. For instance at 2.5:1 ratio, 10 atmos. pressure and at 1550°F the CH_4 in the product is about 8% of the feed. The factors favoring complete CH_4 reaction are high H_2O/CH_4 ratio, high temperature and low pressure. Pressure operation is required to minimize the cost of the reformer unit.

11.8.4 The system shown in Exh. 9-7 operates as follows: The hot reformer gas is cooled and saturated with sulfur vapor in a packed tower over which liquid sulfur is circulated at a 105 gpm. rate. The 900°F temp. rise in the circulating sulfur is handled in the sulfur cooler. This tower section is 4.5 ft. dia. for a superficial velocity of 10' per sec.

11.8.5 The gas (D) leaving the packed section at 300°C carries a small excess of sulfur vapor over that required for the H_2S to be formed. The gas mixture reacts to form H_2S in a two stage catalytic converter with intercooling. The converted gas is cooled to 257°F and is in condition to be mixed with the smelter gas and converter to sulfur in the low temp. Claus system. The H_2S converter beds are 10 ft. dia. and packed with a 2.0 ft. depth of activated alumina.

11.8.6 The quantities shown in Exh. 9-6 produce H_2S equivalent to 70,000 SCFM 2B smelter gas. Stream (L) in Exh. 9-7 is exactly equivalent to stream (K) in Exh. 9-1.

11.8.7 The H_2S from reformed CH_4 is higher in H_2O and is low in inerts as compared to the H_2S from CH_4 -S_2 -H_2O which is lower in H_2O and carries inerts. The equilibrium conversion in the low temperature Claus unit is almost identical on the two H_2S gas sources as shown below.

	Eq. Conv. to Sulfur
Case 1 (CH_4 - S_2 -H_2O)	99.50%
Case 1-b (from reformed CH_4)	99.46%

11.8.8 The economic studies on these directly comparable cases will indicate the lower cost method for H_2S generation to supply a unit producing about 200 tons sulfur per day.

11.8.9 In Exh. 9-8 is presented the results of an equilibrium determination on a reformed methane-sulfur vapor mixture with a feed composition quite close to that used in Case 1-b above.

11. 9 Case 1-c - Same as Case 1-b, except Low Temperature Operation with Cooling by Water Spray Injection

11.9.1 The proposed three stage moving bed reaction system based on external heat exchange, Case 1, seemed a reasonable way to control the reaction temperature. The required service conditions, however, impose very unfavorable operating conditions on these exchangers and they proved to be prohibitively large and expensive. As a result, two alternate systems are proposed which might reduce the cost of heat control on smelter gas.

11.9.2 The first alternate is based on operation at low temperatures (below the melting point of sulfur) and cooling interstage by direct water spray injection. Exh. 9-9 shows this cooling method as applied to the three stage converter. In this system the heat rise is divided equally between three stages operating between 149-241°F. The equilibrium conversion leaving the unit at 241°F is 99.76. The equilibrium constant is so high that the added water has very little effect. It is assumed the process yield and sulfur recovery is identical to that shown in Case 1, Exh. 9-2.

11.9.3 The performance of this system is contingent on the demonstration of favorable kinetics in the lower temperature operating range specified. The advantages are the low cost method of temperature control and good assurance that sulfur vapor will be completely absorbed on the catalyst.

11.10 Case 1-d - Same as Case 1-b, except Single Stage Adiabatic Reactor Based on Vent Gas Recycle

11.10.1 The second alternate reaction system for smelter gas is based on dilution of the smelter gas with a recycle of exit gas sufficient to limit the adiabatic temperature rise to an acceptable maximum. This system is called Case 1-d and it is outlined in Exh. 9-10 and Exh. 9-11.

11.10.2 The converter consists of a moving bed unit 28.6' dia. with a 5.2' deep bed corresponding to a superficial contact time of 1.3 sec. The catalyst inventory in this unit is about 85 tons. As in all previous work on the low temperature Claus system a catalyst loading of 20% sulfur was assumed.

11.10.3 The calculated system yield is 97.6% using vent gas recycle. If air dilution were used, the yield would drop to 95.5%. The use of vent recycle gives the best yield and reduces air pollution to a minimum.

11.10.4 The feasibility of this system depends on the ability of freshly regenerated catalyst to adsorb sulfur vapor completely at 349°F. The counterflow moving bed system will feed cooled regenerated catalyst to the top of the bed and set up the best possible conditions for sulfur vapor adsorption at this higher temperature. A rather comprehensive investigation of sulfur vapor adsorption on the catalyst over a range of temperature would be required in order to judge the merit of the process. The advantage of this system is that the converter is a simple, single stage unit without heat control problems.

11.10.5 A potential advantage is that this is the first case where a fixed bed switch reactor system might be possible. With the three stage unit with intercooling, a switch between conversion and regeneration does not seem practical. If several smaller converters were used instead of the single 28' diameter unit a switch system is at least conceivable. A regeneration system similar to that shown in Exh. 9-5 where the heat is supplied by external hot gas recycle would be required.

11.10.6 The engineering problems in the design of such large units cycling through a wide temperature range might be severe relative to moving bed units operating under steady state conditions. Such a switch system seems dubious based on present information. In any case, it could not be realistically evaluated until the critical requirements in regeneration are defined. In particular, the regeneration cycle must be studied and defined in terms of the minimum time and temperature required, together with the reducing gas requirements in the recycle gas in the unit.

11.10.7 In summary, studies on the low temperature Claus process as applied to smelter gas suggests that there is a reasonable chance of developing a process for this application. With gas above about 2.5% SO_2 a multistage converter with intercooling would be required. A moving bed system seems to be necessary where the multistage reactor is used.

11.10.8 Two alternate moving bed regeneration systems have been considered to work with the moving bed converter.

11.10.9 A single stage adiabatic reactor with the temperature rise limited by vent gas recycle has been proposed. This reactor is the only one which seems to have a potential ability to operate in a conversion-regeneration switch system.

11.10.10 Limiting conditions on both conversion and regeneration must be developed. Sufficient conditions are now available but realistic minimums must be defined before intelligent detailed engineering studies can be made.

12.0 Low Temperature Claus Applied to Power Stack Gas, Type 3

Introduction

12.1 This study was undertaken to obtain a preliminary assessment of the problems which must be faced in adapting the low temperature Claus process to power plant stack gas. The method used is the same as that used on Smelter Gas, namely to visualize a full size unit in some detail and from this define the process and engineering problems as well as possible.

12.2 The principal problems on smelter gas derive from the large adiabatic temperature rise, whereas the problems on power plant stack gas result from the very large volume of weak gas which must be handled.

12.3 Design Problems

Single Fixed Bed Converter

12.3.1 The first approach to the stack gas problem was to consider the converter as a single bed operating at 4 ft./sec. superficial velocity. The No. 3 stack gas condition prescribes 2.371×10^6 SCFM of gas carrying 0.30% SO_2. A converter for this gas condition called for a circular unit 112 ft. dia. or 31 units 20' dia. At 1.35 sec. superficial contact time the bed would be 5 1/2 ft. deep.

12.3.2 A 10 ft./sec. superficial velocity requires a single unit 71 ft. dia. or 13 units 20' dia. with beds 13.5 ft. deep. Each of the conditions meet the SVH value now being used at Princeton Chemical Research in their work on stack gas conversion.

Moving Bed Units

12.3.3 Neither condition appears practical either for a moving bed converter or for a fixed bed switch reactor for the following reasons.

12.3.4 Considering the more favorable unit operating at 10 ft./sec. a single moving bed 71 ft. dia. and 13.5 ft. deep seems an impossible geometry from the standpoint of distributing feed and withdrawing loaded catalyst.

12.3.5 The alternate of 13 moving bed units 20' dia. would lead to a plausible feed and withdrawal system but the multiple units would impose major material handling problems in catalyst handling and gas distribution to maintain balance in all the units.

Multiple Units with Alternate Conversion, Regeneration and Cooling Cycle

12.3.6 The problems in a switch system between conversion and regeneration in the same unit leads to different but equally imposing problems as follows.

12.3.7 If the more favorable 10 ft/sec gas velocity is taken, there will be 13 units on conversion duty.

12.3.8 The catalyst circulation rate at 20% sulfur loading is 215 tons per hour at an overall system yield of 85% and is independent of the type of converter used. The total catalyst inventory is 1350 tons and in a 13 unit system this represents 104 tons catalyst per unit. The 215 ton circulation rate means that a regenerated unit must be put on conversion every half hour.

12.3.9 Such a system with 20 ft. diameter units might result in a process layout as follows:

13 units on conversion on a 6 1/2 hr. cycle
6 " regenerating on a 3 hr. cycle
2 " cooling on a 1 hr. cycle
21 total

12.3.10 A system with units 50 ft. dia. could be represented as follows:

2 units on conversion on a 6 1/2 hr. cycle
1 unit on regeneration and cooling on a 3 1/2 hr. cycle
3 units total

12.3.11 As an initial impression, a switch reactor system based on three 50 ft. dia. units might not seem entirely impractical from the standpoint of tank geometry. There are at least three factors which argue against the switch reactor system as follows.

12.3.12 First, in studies reported later in this report based on a one second contact time and 4/8 mesh Porocel catalyst prohibitive power cost results at the projected 10.0 ft./sec. velocity. The 50 ft. three unit system would have a power cost equivalent to over $300 per ton of sulfur. At a velocity of 4.0 ft./sec. the power cost associated with pressure drop is reduced to an acceptable range. At 4 ft./sec., the three unit system would increase from 50 ft. dia. to 80 ft. or else an equivalent increase in the number of smaller units.

12.3.13 The second and more serious problem in the switch reactor system is supplying the regeneration heat requirements by recycling inert gas, carrying with it a reducing gas such as H_2S needed to avoid catalyst sulfation. If high temperature combustion gas could be used, the duty would be less demanding, but the danger of catalyst overheating and inactivating would be real. The problem of heat loss in large surface area units which supply relatively high temperatures by sensible heat is a significant one in such a regeneration system.

12.3.14 A third problem is the engineering of such large units which must cycle through a wide temperature range without setting up destructive stresses. These 50 to 80 ft. dia. units containing a deep catalyst bed must cycle through a wide temperature range without setting up destructive stresses. These 50 to 80 ft. dia. units containing a deep catalyst bed must cycle through a 600-700°F range every 3 to 6 hours. In a following case we will propose a reactor design based on radial gas flow which may be the basis of a practical design.

12.4 Chain Grate Moving Bed Converter - Case 2

In view of the apparent difficulties in adapting a conventional fixed or moving bed system to the stack gas application, a system based on a catalyst bed moving on a chain grate stoker mechanism was considered first. Refer to Exhibit 9-12.

12.4.1 The regenerated catalyst is distributed on the top chain grate and moves downward from grate to grate. Both gas and catalyst are shown in downward flow so that the gas velocity will not be limited by fluidization or particle elutriation from the bed.

12.4.2 The efficiency of the co-flow unit as shown is not dependent on the use of multiple beds. Fewer beds or even a single bed might be used if mechanical design problems do not prove troublesome. For instance, later we will see that a unit operating at a superficial gas velocity of 10 ft./sec. and a superficial contact time of 1.0 sec. would carry a catalyst inventory of 1000 tons. The cumulative bed depth would be 10.0 ft. which would give 2.5 ft. bed depth on the projected four bed unit.

12.4.3 If pressure drop and power considerations lead to a lower superficial velocity where fluidization and elution would not be a problem, then counterflow would be attractive. In this case, the multiple beds would make a real contribution to efficiency since the freshly regenerated catalyst would be exposed to the lean gas leaving the unit.

12.4.4 Effect of Superficial Velocity on Contact Time

The restricting variable in this design problem is the superficial contact time which relates directly to residence time or SVH. A given contact time can be obtained by different combinations of gas velocity and bed depth. The following table shows the bed area and bed size for a grate unit where the width =0.4 length. All studies were based on 2.35 x 10^6 SCFM #3 Stack Gas.

Table XXII

Superficial Velocity ft./sec.	Bed Area Sq. Ft.	Bed Size Ft.
4	9880	63 x 157
10	3950	40 x 100
20	1970	28 x 70

12.4.5 Other Parameters

In order to indicate how other secondary parameters vary, we have calculated the following table for a superficial velocity of 10.0 ft./sec. Other tables for other velocities could readily be determined.

The following Table XXIII refers to a four bed converter with beds 40 x 100 ft. operating at 10.0 ft./sec., catalyst density, 50#/cu.ft.

Table XXIII

Bed Characteristics vs. SVH

Contact Time Sec.	Cum. Bed Depth-ft.	Bed Depth on each unit - ft.	Catalyst Inventory Tons	Grate Travel ft./min.
0.10	1.0	0.25	100	14.3
0.25	2.5	0.62	250	5.78
0.50	5.0	1.25	500	2.87
1.00	10.0	2.50	1000	1.39
2.00	20.0	5.00	2000	0.72

12.4.6 Power Costs

The very large gas volume together with deep catalyst beds lead to the judgement that the power required to move the gas through the resultant pressure drop will be important. The following tabulation Table 10-1, shows that the power is excessive, i.e. a 4/8 mesh granular catalyst is quite impractical where very large volumes of very dilute gas is being handled. The conditions selected for sizing the chain grate contactor, i.e. 1.0 sec. superficial contact time and 10 ft./sec. superficial gas velocity lead to a ridiculous power cost of $300 per ton of recovered sulfur.

12.4.7 The important factors controlling pressure drop might be altered to give a lower drop as follows:

- The superficial velocity is a major factor since it enters as a square function. This factor may be reduced by increasing the bed area by use of a larger contacting unit or multiple units.

- If experimental work shows that the contact time can be reduced (increased SVH), this would have a direct benefit by reducing the bed depth.

- The use of a larger catalyst size would reduce the pressure drop.

The power consideration is critical on the power stack gas problem where the cost of handling a very large gas volume in a contactor must be applied against a low content of sulfur. Where smelter gas or stronger SO_2 is involved, the power cost per unit of recovered sulfur is not nearly as critical a factor.

12.4.8 It is now apparent that research effort must be directed toward those factors upon which power cost depends before a meaningful evaluation of the application of low temperature Claus to stack gas can be made. While power cost is a less important factor on smelter gas, it will still be an economic factor in the process.

12.4.9 The specific information needed is the relation between SVH and sulfur capture efficiency for a range of catalyst size from that currently in use up to a very large size of perhaps 1 or 2" diameter. Regeneration parameters for the different sizes should be determined. With this information, an approach to process optimization might be made once the factors of gas velocity, bed depth, contactor area and power cost are determined.

12.4.10 As an additional step in defining the scope of the power stack gas problem, the same conditions for a 1.0 inch lump catalyst were calculated, based on the implicit assumption that a volume of 1.0 inch lump catalyst is as effective as the same volume of 4/8 mesh catalyst. It is likely that the large catalyst will not be as effective on a volumetric basis as the smaller size. Any direct comparisons of differential power cost should be made with the realization that the benefit will be reduced by the extent to which bed depth will have to be increased to compensate for lower capture efficiency. This calculation is summarized in the following Table 10-2, to be compared with the results for 4-8 mesh Porocel given in Table 10-1 below.

TABLE 10-1

LOW TEMPERATURE CLAUS - #3 STACK GAS - MOVING BED REACTOR

BASIS: 4-8 MESH POROCEL CATALYST

	Cumulative Bed Depth Ft.			Pressure Drop "H_2O			Horsepower			Power Cost $/ton sulfur		
T \ V >	4	10	20	4	10	20	4	10	20	4	10	20
0.10	0.4	1.0	2.0	6.48	71.5	446.0	5,830	64,250	401,500	2.78	30.07	187.90
0.25	1.0	2.5	5.0	16.2	178.6	1,115	14,570	160,630	1,004,000	6.94	75.17	469.75
0.50	2.0	5.0	10.0	32.4	357.3	2,230	29,150	321,250	2,007,500	13.89	150.34	939.50
1.00	4.0	10.0	20.0	64.8	714.5	4,460	58,300	642,500	4,015,000	27.78	300.69	1,879
2.00	8.0	20.0	40.0	129.6	1,429	8,919	116,600	1,285,000	8,030,000	54.56	601.38	3,758

V - Superficial Velocity - ft./sec.
T - Superficial Contact Time - sec.
Gas permeability based on 4/8 mesh Porocel
60% overall electrical and mechanical efficiency on blower
900 HP/"H_2O pressure drop
HP x 0.000468 = $/ton sulfur
2.37×10^6 SCFM = 3.43×10^6 CFM at 257°F
257°F bed temperature
384.7 tons sulfur/day produced
Electric power $0.0075/HP hour

TABLE 10-2

LOW TEMPERATURE CLAUS - #3 STACK GAS - MOVING BED REACTOR

BASIS: 1" LUMP CATALYST

	Cumulative Bed Depth Ft.			Pressure Drop "H_2O			Horsepower			Power Cost $/ton sulfur		
T ↓ → V	4	10	20	4	10	20	4	10	20	4	10	20
0.10	0.4	1.0	2.0	0.51	5.42	34.2	460	4,880	30,780	0.22	2.28	14.41
0.25	1.0	2.5	5.0	1.27	13.50	85.5	1,140	12,180	76,950	0.54	5.69	36.01
0.50	2.0	5.0	10.0	2.54	27.10	171.0	2,280	24,390	153,900	1.07	11.38	72.03
1.00	4.0	10.0	20.0	5.08	54.20	342.0	4,560	48,780	307,800	2.14	22.76	144.05
2.00	8.0	20.0	40.0	10.16	108.40	684.0	9,120	97,560	615,600	4.28	45.52	288.10

V = Superficial velocity - ft./sec.
T = Superficial contact time - sec.
Gas permeability based on 1" lump catalyst
60% overall electrical and mechanical efficiency on blower
900 HP/"H_2O pressure drop
HP x 0.000468 = $/ton sulfur
2.37 x 106 SCFM = 3.43 x 106 CFM at 257°F
257°F bed temperature
384.7 tons sulfur/day produced
Electric power $0.0075/HP hour

Visualized Process Layout

12.4.11 As a final step in the consideration of the stack gas problem, a process layout was projected for a system using the chain grate type converter, H_2S generated from reformed methane and regeneration in an indirectly heated unit using condensing sulfur vapor as the heating agent. The layout is carried in Exh. 9-12 and the corresponding process profile is carried in Exh. 9-13. Process details were submitted to Engineering, for a capital estimate on the information covered in these exhibits.

12.4.12 Process conditions as follows were established.

- The H_2S system is based on reformed methane. The plant represents a 2.24 scale factor over the design for Case 1-B.
- The regenerator represents a 2.15 scale up from the Case I system.
- The chain grate converter is based on a superficial velocity of 10 ft./sec. leading to a grate size of 40 x 100 ft.
- Pressure drop is based on 1.0 inch lump catalyst. This would require a wire mesh traveling with the grate.
- Catalyst circulation rate 215 tons/hr.
- Catalyst loading 20% sulfur
- Catalyst inventory 1000 tons

12.4.13 The implications of these preliminary studies seem clear and may be summarized as follows:

- A catalyst granule substantially larger than the 4/8 mesh size first considered must be used.
- The 10 ft./sec. gas velocity is unrealistic and the bed area must be increased to give a more moderate velocity.
- The minimum contact time must be established for each catalyst size and used in the process visualizations.

12.5 Case 2B - Power Stack Gas, Type 3 (Switch Reactors)

Introduction

12.5.1 In the case 2R preceding there was outlined the application of a moving bed system as applied to stack gas. Such a process is open to serious problems in the transport of catalyst. The quantity in circulation is large and will be subject to attrition losses of undefined magnitude. If a fixed bed system could be adapted to the stack gas problem, the benefits would be very great. No catalyst transport problems occur and catalyst loss due to attrition should be negligible. In such a system, catalyst would be replaced due to loss in activity rather than by loss from the system as dust.

12.5.2 The problems relative to a fixed bed-switch reactor system as applied to stack gas have been discussed under Case 2R. These problems influenced the adoption of a moving bed traveling on a chain grate as an initial process approach. In this instance, a detailed study of a fixed bed system is supplied which will provide a comparison with the moving bed system.

Process Considerations - Switch Reactor System

12.5.3 The reaction of H_2S and SO_2 to give sulfur is not a problem in a fixed bed. This sulfur forming step proceeds with equal ease in either the fixed or moving bed system.

12.5.4 The regeneration stage is where the major design and operation problems arise. These problems relate mainly to the necessity for supplying regeneration heat as sensible heat in the entering gas, plus the valving and manifolding required to direct the very large volumes of stack gas and regeneration gas to the units operating in a sequential way. An overall description of the operation of the system is as follows. (Refer to Exh. 11-8)

12.5.5 H_2S is generated by the reaction of reformer gas and sulfur vapor in a catalytic reactor. The incoming flue gas is cooled to the reactor inlet temperature by water spray in a spray chamber. The cooled flue gas is mixed with H_2S in the proper ratio and passed to two low temperature Claus reactors in parallel. The gas leaving the reactors is indicated as a vent, although it carries an unacceptable content of H_2S.

12.5.6 The two low temperature Claus reactors operate in parallel on conversion on a 10 hour cycle. The converter operation is staggered so that every 5 hours a sulfur loaded unit is switched to regeneration and a cooled regenerated unit is switched to conversion.

12.5.7 On regeneration mode the heat is supplied to the loaded unit by a recycling hot gas. The loaded unit and the hot newly regenerated unit are in series so that the recycle gas will cool the regenerated unit and at the same recuperate heat.

12.5.8 By reference to Exh. 11-8, the cool gas leaving the sulfur condensers is circulated by a blower thru the hot regenerated unit where it picks up heat. The gas is then further heated in the gas reheat furnace and passed into the unit to be regenerated. The gas leaving the unit on regeneration is cooled in the sulfur condenser, the sulfur is separated and the gas is recycled. The medium pressure recovered steam is generated in this condenser system.

12.5.9 A serious problem may occur when a cooled regenerated reactor is switched to conversion duty. In this condition the regenerated activated alumina catalyst will act as a dessicant and pick up a large amount of water from the stack gas carrying over 10% H_2O. The heat of adsorption will give a large heat kick of undefined magnitude. The implications of this water adsorption are difficult to appraise. As the catalyst loads with water, a hot zone will travel thru the bed. Continuing flow of stack gas will cool the bed and the water loaded catalyst should perform in a normal way in sulfur conversion.

12.5.10 The minimum penalty of the temperature excursion will be a loss of cycle time at optimum conditions. The most drastic result would be the loss of the sulfur formed during the initial period as sulfur vapor to the stack when the hot zone traverses the bed.

12.5.11 The adsorbed water may be desorbed later in the cycle as sulfur accumulates on the catalyst. One would anticipate at least a partial desorption of water but not a total desorption.

12.5.12 When a sulfur loaded catalyst carrying adsorbed water is switched to regeneration, the water will be released as vapor into the gas recycle system. It is calculated that if the converter unit going on regeneration carries 1.6% water by weight the volume of steam released will equal the volume of the system on regeneration and cooling. The large volume of steam released means that the system must be vented to avoid a pressure rise and this will carry the inert gas out of the system. This venting is complicated by the necessity of maintaining a reducing gas such as H_2S in the recycle system in order to maintain catalyst activity.

12.5.13 Probably the most practical solution would be to adopt superheated steam instead of nitrogen as the inert gas in the regeneration system. In any case, taking a small flow of H_2S from the H_2S generator into the regeneration system is visualized. The regeneration system would vent into the flue gas system at the main blower inlet. In this way the H_2S content in the regeneration system vent would be utilized in the sulfur forming reaction.

12.5.14 In the following process study, it is assumed that the heat of water adsorption will cause no process difficulties. Also ignored is any steam release on regeneration and it is assumed the inert gas is nitrogen.

12.5.15 Another special consideration deriving from the switch reactor system is the question of steam credits. The position is taken that only a steady flow of recovered steam may receive a valid credit. The recovery steam flow in the system as outlined would be highly variable through the cycle ranging from a very low value at switch over and then thru a variable rate to some maximum. There would be no period of steady flow at an average value.

12.5.16 The following discussion will point up the problem. At switch over the following shifts are made.

- A loaded converter is moved to regeneration.
- A regenerated unit is moved to cooling.
- A cooled regenerated unit is moved to conversion.

A half loaded unit remains on conversion.

12.5.17 After this shift the regenerator system starts with a maximum preheat obtaining on the recycle gas and consequently, a minimum demand on the regenerator gas reheater. Light firing in the reheater causes a small rate of high pressure steam from the heat recovery system in the flue gas from the reheater. As the unit on cooling loses heat, the firing rate on the regenerator reheater is increased to maintain the fixed gas temperature entering the unit on regeneration and the generation of high pressure steam increases.

12.5.18 In another way the intermediate pressure steam generated in the sulfur recovery gas cooler system is variable. At the start of regeneration the cold catalyst bed picks up substantially all the heat and steam recovery from the gas coolers is very small. As the bed in the regenerator heats up the gas temperature leaving increases and the steam generation steadily increases to a maximum at the end of the regeneration cycle. Our calculations indicate that 79% of the heat is used to heat the catalyst while only 21% is required to heat and vaporize the sulfur. The large heat capacity of the catalyst bed will lead to a gradual increase in gas outlet temperature over the five hour regeneration cycle.

12.5.19 In this study, it is attempted to maximize the quantity of steam for credit as follows. During the early portion of the regeneration cycle the medium pressure steam from the sulfur condenser is augmented or totally supplied by high pressure steam from the regenerator gas reheater. The high pressure steam is let down to the intermediate steam pressure as required. During periods of light duty on the gas reheater, the steam flow is augmented by firing below the economizer section of the reheater in order to generate the required high pressure steam.

12.5.20 In this way, the medium pressure steam is continuous and can be used for turbine power despite the cyclic nature of regeneration temperatures. The required supplemental fuel is included in the fuel gas requirement for the system. This fuel is well justified by the steam credit made available.

Process Projection - Switch Reactor System

12.5.21 The flow sheet for the system is carried in Exh. 11-18 and the process profile is carried in Exh 11-9. The values for flows and temperatures shown in the profile represent average values over the complete cycle. These average values represent close material and energy balances. The complex cycling of instant values would be very difficult to estimate and were not considered essential to this study.

12.5.22 A radial flow converter with catalyst in an annular ring was developed which seems to overcome the potential stress problems in the alternate heating and cooling of a conventional reactor carrying horizontal beds. This unit is shown as Exh. 11-10.

12.5.23 Fixed System Parameters

2.0 sec. superficial contact time on conversion
4.0 ft./sec. superficial velocity on conversion
20.0% sulfur loading on catalyst
1.0 inch lump activated alumina catalyst
8.0 ft. catalyst bed depth in radial flow reactor
Gas into converter at 225°F and out at 248°F
Stack gas cooling with spray water.

12.5.24 Derived System Parameters

10.0 in. H_2O press. drop on converter bed
0.1 in. H_2O press. drop on regenerator bed
5145 sq. ft. bed area in each unit
41160 cu. ft. catalyst in each unit
1030 tons catalyst in each unit
4120 tons catalyst in the system
9350 horsepower on stack gas blower at 60% Eff. for bed drop only.
20,000 horsepower total plant requirement.
10.0 hrs. conversion cycle.
5.0 hrs. regeneration cycle
5.0 hrs. cooling cycle

Average duty regenerator gas reheater	1.76×10^6 BTU/min.
Average duty condenser cooler	1.19×10^6 BTU/min.
Spray water to flue gas cooler	525 gpm

12.5.25 This study has indicated two problems which would require experimental study if a switch reactor system such as this were to be considered.

- The extent of water adsorption and the implications of this adsorption on the process should be determined.

- The regeneration step using pure steam or a recycle gas high in water vapor should be studied. The conditions necessary to maintain catalytic activity using such a recycle gas should be demonstrated.

12.6 Case 2A - Moving Bed Low Temperature Claus Applied to No. 3 Stack Gas

12.6.1 In this case a moving bed alternate to chain grate system used in case 2R is presented as a less costly system. The proposed reactor concept is shown in Exhibit 11-13. In this reactor the stack gas enters between two rectangular beds. The gas flows across the beds and is directed to the stack. The beds are vertical and rectangular with the regenerated catalyst feed at the top. The catalyst flows downward at a regulated rate governed by the bottom discharge mechanism. The indicated screw conveyor may be less desirable than a flow regulated by a moving belt at the discharge.

12.6.2 The reactor has been sized for a 4 ft./sec. superficial velocity and a 2 sec. superficial contact time. These are the basic parameters used in the switch reactor Case 2B and give a low pressure drop across the bed.

12.6.3 All the conditions carried in the process profile for Case 2B are maintained. 20% sulfur loading is assumed and the regeneration section described for Case 2B is identical. The utilities are the same as for Case 2B except for 15,000 horsepower less as electricity. The capital requirement for this alternate system was obtained and the process economics were worked up, as shown in Exhibits 11-11 and 11-12.

12.7 Comment on Sulfur Loading

12.7.1 All the cases projected covering aspects of the Low Temperature Claus Process have assumed a 20% sulfur loading on the catalyst. The process implications of a failure to attain this loading are serious. Both the moving bed system and the switch reactor system would be adversely affected for a lower loading but in different ways.

12.7.2 In the switch bed system at 20% sulfur loading the heat supplied to the regeneration is distributed as 79% to heat the catalyst and 21% to vaporize the sulfur. At a 10% loading the heat demand would increase by a factor of about 1.7 relative to the 20% loading. The cycle time would be cut in half and it is apparent that the maximum heat flux would be sharply increased. In this cyclic system the regeneration gas reheater and the sulfur condensers must be sized for the peak load.

12.7.3 A second effect would be to cut the average sulfur vapor level in the regenerator gas from 1.1% on a 20% loading at about 0.65% on a 10% loading and thus make the sulfur more difficult to recover.

12.7.4 On the moving bed systems a drop in sulfur loading from 20% to 10% would double the catalyst recirculation rate and would double the catalyst loss due to attrition.

12.7.5 There is not sufficient information available to estimate the sulfur loading at which these stack gas systems become inoperative or prohibitively costly. It is certain however, that a loading substantially below 20% would adversely affect the economic picture projected for 20% sulfur loading.

12.8 Case 6 - Low Temperature Claus Applied to No. 2C Smelter Gas

12.8.1 Process studies directed toward this 2C smelter gas are considered particularly significant. This gas carries 2.9% SO_2 and is the weakest gas we are considering. If this gas can be economically processed, it would mean that virtually the entire range of smelter gas would be amenable to recovery.

12.8.2 The converter system is similar to that used on the 2B smelter gas Case 1-C with certain modifications. The lower SO_2 content in the 2-C gas allows the temperature rise to be controlled in a two-zone converter rather than three. Direct spray water injection is used for precooling and intercooling, as in Case 1-C.

12.8.3 The converter is designed on the following basis:

Superficial contact time	2.0 sec.
Superficial velocity	4.5 ft./sec.
Sulfur loading	20%

12.8.4 This works out to a unit 32.5 ft. dia. with two beds each 4.5 ft. deep. The catalyst inventory is estimated at 191 tons. The process layout is shown in Exhibit 10-13 and the detailed process profile is given in Exhibit 10-14. The H_2S unit is derived from reformed methane and represents a 1.93 scale-up from the original unit shown for Case 1b. The regenerator is based on indirect heating with sulfur vapor and represents a 1.95 scale-up from the original unit shown for Case 1.

12.8.5 The catalyst cooler is based on the use of 77°F air counterflow to the catalyst. An air flow of about 60,000 SCFM would be required.

12.8.6 With respect to pressure drop and power cost, this 2-C gas is the less favorable of the smelter gas cases due to the low SO_2 content. We have selected this case as the basis of a more comprehensive look at the relation between catalyst particle diameter and the power cost equivalent to pressure drop.

12.8.7 The conditions used in the calculations are as follows:

Sulfur production	385 tons/day
Superficial velocity	4.5 ft./sec.
Mean bed temperature	195°F
Bed depth	9.0 ft.
Gas Volume	219,300 SCFM
Gas temp. at blower	150°F
60% overall blower efficiency	
Power cost	$0.01/KWH

12.8.8 The following tabulation summarizes the result of pressure drop calculations on a series of different catalyst forms. It is again emphasized that the power cost is based on the assumption that all the catalysts operate at the same SVH.

Table XXIV

Catalyst Size	Form	Bed ΔP Inch H_2O	Horsepower	Power Cost $/ton sulfur
4/8 mesh	Granular	162	12,800	6.00
1/4 inch	Granular	81	6,400	3.01
1/2 inch	Granular	31	2,440	1.15
3/4 inch	Granular	18	1,420	0.67
1.0 inch	Granular	12	940	0.44
5/16 Porocel	LPD	52	4,100	1.92
5/8 Porocel	LPD	20	1,560	0.73

12.8.9 The tabulation suggests that a reasonable power cost is attainable on 2C smelter gas at the contact time and gas velocity assumed in the design. If the gas were split between two reactors of the design indicated, the power cost would be cut to about 1/4 of the values indicated above. When performance data on the different catalyst forms is available, the balance between reactor cost and power cost can be optimized.

13.0 Reduction of SO_2 with Metallic Sulfides

13.1 We have been requested to look into the reaction of SO_2 with metallic sulfides. The object is to define the feasibility of sulfur recovery by reaction with SO_2.

13.2 The equilibrium conditions applying to common sulfide ores were calculated according to the following reactions:

(1) $3FeS + 2SO_2 \longrightarrow Fe_3O_4 + 5/2\ S_2$
(2) $2ZnS + SO_2 \longrightarrow 2ZnO + 3/2\ S_2$
(3) $2PbS + SO_2 \longrightarrow 2PbO + 3/2\ S_2$
(4) $2Cu_2S + SO_2 \longrightarrow 2Cu_2O + 3/2\ S_2$

13.3 The attached plot, Exhibit 6-22, shows the results of the calculations expressed as Log K vs. Temp. (°K). This plot shows the FeS reaction to be the most favorable and that 1200 K is the most favorable temperature. The three crosses plotted are the results of calculations made by a Russian group and reported in (CA 67, 34508 V).

13.4 Curve 3 was extrapolated to 1200°K as shown and the equilibrium gas composition at 1200°K was calculated for the four reactions. The calculations are summarized as follows:

TABLE XXVI

Gas Composition at 1200° Kelvin (1700°F)

Reaction	Log K	K	% S_2 in Eq. Gas	% SO_2 Converted
1	-3.60	2.5×10^{-4}	1.46	1.17
2	-5.00	1.0×10^{-5}	6.7×10^{-2}	4.5×10^{-2}
3	-7.75	1.78×10^{-8}	7.0×10^{-4}	4.7×10^{-4}
4	-12.26	5.5×10^{-13}	6.0×10^{-7}	4.0×10^{-7}

13.5 The above equilibrium compositions represent the $\%S_2$ in the SO_2 gas at equilibrium at 1200°K. The gas is in equilibrium with solid phases of both sulfide and oxide.

13.6 The extremely low conversion per pass makes this route entirely impractical as applied to these common sulfide minerals.

14.0 EQUILIBRIUM CALCULATIONS - THERMODYNAMIC BACKGROUND

14.1 This section is presented in order to describe in some detail the thermodynamic basis for the equilibrium calculations performed in the course of our Phase I studies.

14.2 The principal problem in defining the equilibrium conditions in a complex system such as the reduction of SO_2 with methane lies in the selection of a consistent set of independent equations. Once a suitable set of equations has been selected the organization of the equilibrium relations is routine.

14.3 H. J. Kandiner and S. R. Brinkley, Jr. of the U. S. Bureau of Mines has defined a systematic approach to this problem which leads to the selection of a suitable set of equations without ambiguity. This selection procedure is detailed in Ind. & Eng. Chem., May 1950, pp. 850-855. "Calculation of Complex Equilibrium Relations". In the course of the Phase I studies we have derived equilibrium data for three different situations, as follows.

14.4 The reaction of SO_2 with CH_4. Depending on input ratios and water content, this system may lead to primary production of S_2, S_6, S_8 or H_2S or CS_2 or other components in the system. The same equilibrium system may be used for the reduction of SO_2 with H_2 or CO or mixtures of H_2 and CO.

14.5 The reaction of SO_2 with solid carbon. When solid carbon is present a different treatment is required.

14.6 The reaction of H_2S and SO_2 at low temperatures on a catalyst under conditions where liquid or solid sulfur is adsorbed by the catalyst. The special situation is to determine equilibrium conversion when the sulfur vapor in the gas phase is determined by the vapor pressure of the condensed sulfur phase.

14.7 The second part of the Kandiner and Brinkley paper describe a system for the numerical solution of the equilibrium expressions.

14.8 This procedure would be practical for the solution of a few specific equilibrium conditions but would be prohibitively time consuming by manual calculations for the systematic study of many input compositions over a wide temperature range at narrow temperature intervals.

14.9 Allied Chemical has developed a computer program for equilibrium calculations which we have used as required in developing a sound thermodynamic reference background for the Phase I studies. The comprehensive equilibrium studies would not have been possible within the Phase I scope without computer solution.

14.10 Reduction with Methane

14.10.1 The derivation of the equilibrium system for the methane reduction of SO_2 will be described in detail. It is recommended that the Kandiner and Brinkley paper should be in hand for reference.

14.10.2 The first step is to set up a table having for column headings the various chemical elements in the system and a listing of the chemical species in the left column. The entries are the subscripts to each element in the various chemical formulas of the species. For the methane reduction system this table is as follows:

TABLE XXVII

	S	O	C	H
SO_2	1	2	0	0
CH_4	0	0	1	4
CO_2	0	2	1	0
CO	0	1	1	0
H_2O	0	1	0	2
H_2S	1	0	0	2
COS	1	1	1	0
CS_2	2	0	1	0
S_2	2	0	0	0
S_6	6	0	0	0
S_8	8	0	0	0
H_2	0	0	0	2

14.10.3 The selection procedure states that if a non-zero fourth order determinant can be found using the four columns and any four rows of stoichiometrically independent compounds (S_2, S_6 & S_8 are not independent), then there will be exactly four independent components in the system and equations must be written for the formation of the other components.

14.10.4 In this system several non-zero fourth order determinants can be formed such as the following:

$$\begin{array}{lr} SO_2 & 1200\ ) \\ CH_4 & 0014\ ) \\ H_2S & 1002\ ) \\ CO & 0110\ ) \end{array} \quad = \quad -4$$

$$\begin{array}{lr} SO_2 & 1200\ ) \\ CH_4 & 0014\ ) \\ CO_2 & 0210\ ) \\ COS & 1110\ ) \end{array} \quad = \quad -8$$

H_2O	0102)	
CO_2	0210)	= -4
H_2S	1002)	
S_2	2000)	

14.10.5 In the system there are four independent components and since there are twelve components involved we must set up equations for the other eight in terms of the dependent components.

14.10.6 The following set of equations serve to define the system. The four independent components can be freely selected. Once selected, the dependent equations are unique for the set of independent components selected. The independent components selected are H_2O, S_2, H_2 and CO. The equations are numbered and these numbers will be used to identify the equation in later references. The expression for the equilibrium constant for each equation is shown for each reaction.

Eq. No.		For
1	$4H_2O + S_2 \longrightarrow 4H_2 + 2SO_2$	SO_2
2	$CO + 3H_2 \longrightarrow H_2O + CH_4$	CH_4
3	$2H_2 + S_2 \longrightarrow 2H_2S$	H_2S
4	$CO + H_2O \longrightarrow H_2 + CO_2$	CO_2
5	$2CO + S_2 \longrightarrow 2\ COS$	COS
6	$S_2 + CO + H_2 \longrightarrow H_2O + CS$	CS_2
7	$3S_2 \longrightarrow S_6$	S_6
8	$4S_2 \longrightarrow S_8$	S_8

$$K_1 = \frac{(H_2)^4\ (SO_2)^2}{(H_2O)^4\ (S_2)}$$

$$K_2 = \frac{(H_2O)\ (CH_4)}{(CO)\ (H_2)^3}$$

$$K_3 = \frac{(H_2S)^2}{(H_2)^2\ (S_2)}$$

$$K_4 = \frac{(H_2)\ (CO_2)}{(CO)\ (H_2O)}$$

$$K_5 = \frac{(COS)^2}{(CO)^2 \quad (S_2)}$$

$$K_6 = \frac{(H_2O) \quad (CS_2)}{(S_2) \ (CO) \ (H_2)}$$

$$K_7 = \frac{(S_6)}{(S_2)^3}$$

$$K_8 = \frac{(S_8)}{(S_2)^4}$$

14.10.7 The next step is to set up the material balance for the system. In this balance we have shown SO_2, CH_4, CO_2 and H_2O as inlet moles. It is clear that any other components could be included if desired. In our computer program we can take any combination of the twelve components as inlet moles. In this system as defined here there must be 100 moles of feed gas so that the individual components enter as mole %. The material balance relations are as follows:

Basis 100 moles inlet gas

Total pressure 1.0 atmosphere

S = moles S_2 reacted by (1)
T = " CO " " (2)
U = " S_2 " " (3)
V = " CO " " (4)
W = " CO " " (5)
X = " CO " " (6)
Y = " S_2 " " (7)
Z = " S_2 " " (8)

Inlet Moles
A = SO_2 moles
B = CH_4 "
C = CO_2 "
D = H_2O "

$$SO_2 = A + 2S$$
$$CH_4 = B + T$$
$$CO_2 = C + V$$
$$CO = -T - V - W - X$$
$$H_2O = D - 4S + T - V + X$$
$$S_2 = -S - U - \frac{W}{2} - X - Y - Z$$

$$S_6 = \frac{Y}{3}$$

$$S_8 = \frac{Z}{4}$$

$$H_2S = 2U$$

$$COS = W$$

$$CS_2 = X$$

$$H_2 = 4S - 3T - 2U + V - X$$

$$\text{Total Moles} = \left(100 + S - 2T - U - \frac{W}{2} - X - \frac{2Y}{3} - \frac{3A}{4}\right) = R$$

14.10.8 The final step is to substitute the expressions from the material balance into the equilibrium relations as expressed for each reaction. The final equilibrium relations are as follows:

$$K_1 = \frac{(4S-3T-2U+V-X)^4 (A+2S)^2}{(D-4S+T-V+X)^4 \left(-S-U-\frac{W}{2}-X-Y-Z\right)(R)}$$

$$K_2 = \frac{(D-4S+T-V+X)(B+T)(R)^2}{(-T-V-W-X)(4S-3T-2U+V-X)^3}$$

$$K_3 = \frac{(2U)^2\ (R)}{(4S-3T-2U+V-X)^2 \left(-S-U-\frac{W}{2}-X-Y-Z\right)}$$

$$K_4 = \frac{(4S-3T-2U+V-X)(C+V)}{(-T-V-W-X)(D-4S+T-V+X)}$$

$$K_5 = \frac{(W)^2 (R)}{(-T-V-W-X)^2 \left(-S-U-\frac{W}{2}-X-Y-Z\right)}$$

$$K_6 = \frac{(D-43+T-V+X)(X)(R)}{\left(-S-U-\frac{W}{2}-X-Y-Z\right)(-T-V-W-X)(4S-3T-2U+V-X)}$$

$$K_7 = \frac{\left(\frac{Y}{3}\right)\ (R)^2}{\left(-S-U-\frac{W}{2}-X-Y-Z\right)^3}$$

$$K_8 = \frac{\left(\frac{Z}{4}\right)\ (R)^3}{\left(-S-U-\frac{W}{2}-X-Y-Z\right)^4}$$

14.10.9 The value of the equilibrium constant is evaluated for each of the reactions at the temperature of interest. The quantities S, T, U, V, W, X, Y and Z are determined by the solution of the eight simultaneous equations. Substitution of these determined values back into the material balance gives the equilibrium moles desired for the given input composition and temperature.

14.10.10 Free energy values for evaluation of the equilibrium constants is taken from the JANAF tables with the single exception of S_6 for which there is no JANAF table. Free energy for S_6 is taken from the "Sulfur Data Book" published by the Freeport Sulfur Co. (1954 Ed). ΔF for the reaction $6\ S_{Rh} \longrightarrow S_6$ is given by the following equation, ΔF in gm Cal/gr formula mass, T in °K

$$\Delta F = 29990 + 5.112T \log T + 0.0174\ T^2 - 71.60\ T$$

14.10.11 The equilibrium program for the methane system just described is more versatile than would first appear. As far as we know any equilibrium involving part or all of the twelve components of the system may be determined. The only limitation is that the four elements C, O, H and S must be present or the computer will be unable to come to a solution. For instance, this program was used for the reduction of SO_2 with H_2. In this case there is no carbon in the feed and the computer fails in the attempt to match equilibrium constants involving carbon containing species. This condition is readily handled by adding a trace of carbon containing compound to the feed composition. An example of this would be as follows:

TABLE XXVIII

	Original feed	Modified feed
H_2	20.00	20.00
SO_2	10.00	10.00
N_2	70.00	69.99
CO_2		0.01

14.10.12 In a similar way if one were determining the equilibrium in the synthesis of CS_2 from methane and sulfur vapor according to the principal reaction

$$2\ S_2 + CH_4 \longrightarrow CS_2 + H_2S$$

0.01 moles H_2O would be added to the 100 moles of feed to allow matching the equilibrium constants involving oxygen bearing compounds.

14.10.13 In the equilibrium studies the I.B.M. 360 computer is operated in the double **precision** mode and it was found that the addition of 0.01 mole of a compound carrying the missing element has proven sufficient. This small addition, of course, has a negligible effect on the equilibrium gas composition.

14.11 Reduction with Carbon

14.11.1 The equilibrium system for solid carbon reduction is set up in the same way as that for CH_4 reduction. There is an added restriction where a solid phase exists which requires that this phase be selected as one of the independent components.

14.11.2 A table is set up from the elements and chemical species as before. This table is as follows:

TABLE XXIX

	S	O	H	C
SO_2	1	2	0	0
H_2O	0	1	2	0
CO_2	0	2	0	1
CO	0	1	0	1
H_2	0	0	2	0
S_2	2	0	0	0
H_2S	1	0	2	0
COS	1	1	0	1
CS_2	2	0	0	1
S_6	6	0	0	0
S_8	8	0	0	0
C	0	0	0	1

14.11.3 A non zero fourth order determinant can be formed as follows giving us four independent components:

$$\left.\begin{array}{lcccc} SO_2 & 1 & 2 & 0 & 0 \\ H_2 & 0 & 0 & 2 & 0 \\ S_2 & 2 & 0 & 0 & 0 \\ C & 0 & 0 & 0 & 1 \end{array}\right) = -4$$

We will select these four species as the independent components.

14.11.4 The following set of equations is derived to express the eight dependent components in terms of the independent components.

		For
1	$H_2 + SO_2 \longrightarrow 1/2\ S_2 + 2H_2O$	H_2O
2	$C + SO_2 \longrightarrow 1/2\ S_2 + CO_2$	CO_2
3	$2C + SO_2 \longrightarrow 1/2\ S_2 + 2CO$	CO
4	$2H_2 + S_2 \longrightarrow 2H_2S$	H_2S
5	$2C + SO_2 + 1/2\ S_2 \longrightarrow 2\ COS$	COS
6	$C + S_2 \longrightarrow CS_2$	CS_2
7	$3S_2 \longrightarrow S_6$	S_6
8	$4S_2 \longrightarrow S_8$	S_8

14.11.5 Equilibrium expressions corresponding to the above set of equations is as follows:

$$K_1 = \frac{(S_2)^{1/2}(H_2O)^2}{(H_2)(SO_2)}$$

$$K_2 = \frac{(S_2)^{1/2}(CO_2)}{(SO_2)}$$

$$K_3 = \frac{(S_2)^{1/2}(CO)^2}{(SO_2)}$$

$$K_4 = \frac{(H_2S)^2}{(H_2)^2\ (S_2)}$$

$$K_5 = \frac{(COS)^2}{(SO_2)(S_2)^{1/2}}$$

$$K_6 = \frac{(CS_2)}{(S_2)}$$

$$K_7 = \frac{(S_6)}{(S_2)^3}$$

$$K_8 = \frac{(S_8)}{(S_2)^4}$$

14.11.6 The material balance is set up using the following symbols:

S = moles SO_2 reacted by (1)
T = " " " " (2)
U = " " " " (3)
V = " S_2 " " (4)
W = " SO_2 " " (5)
X = " S_2 " " (6)
Y = " S_2 " " (7)
Z = " S_2 " " (8)

Inlet

A = SO_2 moles
B = H_2O "
C = CO_2 "
D = CO "

Basis 100 moles inlet gas

SO_2 = A-S-T-U-W
H_2O = B + 2S
CO_2 = C + T
CO = D + 2U
H_2 = -S -2V
H_2S = 2V
COS = 2W
CS_2 = X
$S_2 = \frac{S}{2} + \frac{T}{2} + \frac{U}{2} - V - \frac{W}{2} - X - Y - Z$
S_6 = Y/3
S_8 = Z/4
Total Moles = R = $\left(100 + \frac{S}{2} + \frac{T}{2} + \frac{3U}{2} - V + \frac{W}{2} - \frac{2Y}{3} - \frac{3Z}{4}\right)$

14.11.7 The following set of simultaneous equations is solved for S,T,U,V,W,X,Y and Z for desired temperatures and gas inlet compositions. These values are substituted into the material balance to give the equilibrium moles.

$$K_1 = \frac{\left(\frac{S}{2} + \frac{T}{2} + \frac{U}{2} - V - \frac{W}{2} - X - Y - Z\right)^{1/2} (B+2S)^2}{(-S-2V)(A-S-T-U-W)\ (R)^{1/2}}$$

$$K_2 = \frac{\left\{\frac{S}{2} + \frac{T}{2} + \frac{U}{2} - V - \frac{W}{2} - X - Y - Z\right\}^{1/2} (C+T)}{(A-S-T-U-W)(R)^{1/2}}$$

$$K_3 = \frac{\left\{\frac{S}{2} + \frac{T}{2} + \frac{U}{2} - V - \frac{W}{2} - X - Y - Z\right\}^{1/2} \left\{D + 2U\right\}^2}{(A-S-T-U-W)\ (R)^{3/2}}$$

$$K_4 = \frac{(2V)^2\ (R)}{(-S-2V)^2 \left\{\frac{S}{2} + \frac{T}{2} + \frac{U}{2} - V - \frac{W}{2} - X - Y - Z\right\}}$$

$$K_5 = \frac{(2W)^2}{(A-S-T-U-W)\left\{\frac{S}{2} + \frac{T}{2} + \frac{U}{2} - V - \frac{W}{2} - X - Y - Z\right\}^{1/2}(R)^{1/2}}$$

$$K_6 = \frac{(X)}{\left\{\frac{S}{2} + \frac{T}{2} + \frac{U}{2} - V - \frac{W}{2} - X - Y - Z\right\}}$$

$$K_7 = \frac{(Y/3)\ (R)^2}{\left\{\frac{S}{2} + \frac{T}{2} + \frac{U}{2} - V - \frac{W}{2} - X - Y - Z\right\}^3}$$

$$K_8 = \frac{(Z/4)\ (R)^3}{\left\{\frac{S}{2} + \frac{T}{2} + \frac{U}{2} - V - \frac{W}{2} - X - Y - Z\right\}^4}$$

14.12 Low Temperature Claus Equilibrium Studies

14.12.1 At low SO_2 levels such as in stack gas or very dilute smelter gas, calculations show that a serious sulfur loss would occur if sulfur vapor were in equilibrium with the solid or liquid sulfur phase. To avoid this loss, the catalyst must exert a strong adsorptive action on the sulfur vapor.

14.12.2 The purpose of the equilibrium calculations was to separate the effects of catalytic conversion and sulfur vapor adsorption and to show how each affects the recoverable sulfur.

14.12.3 The equilibrium calculations are unusual since, where equilibrium between sulfur and its vapor is assumed, the S_8 fraction in the gas is determined by sulfur vapor pressure and not by material balance. S_6 will be present to a small extent in the higher temperature range but the vapor was assumed to be entirely S_8.

14.12.4 This work was not programmed for computer solution since it was a special one-time study which was quite practical in manual computation.

14.12.5 The first step was to calculate the vapor pressure of sulfur at temperatures in the range of interest. The following equations were used.

For solid monoclinic sulfur (100°C)

$$\text{Log } P = 11.364 - \frac{5082}{T}$$

For liquid sulfur (120-200°C)

$$\text{Log } P = 14.700 - 0.006224\ T - \frac{5405.1}{T}$$

P - millimeters Hg

T - °K

Vapor pressures are calculated as follows.

TABLE XXX

°C	°K	Log P	P mm.	Mol % S_8	8(%S_8)	°F
100	373	-2.261	5.48×10^{-3}	0.00072	0.0058	212
120	393	-1.4994	3.16×10^{-2}	0.0042	0.034	248
125	398	-1.3577	4.38×10^{-2}	0.0058	0.046	257
140	413	-0.9578	1.10×10^{-1}	0.0145	0.116	284
160	433	-0.4778	3.32×10^{-1}	0.043	0.345	320
170	443	-0.2586	5.50×10^{-1}	0.072	0.576	338
180	453	-0.0512	8.90×10^{-1}	0.131	1.050	356
200	473	+0.3288	2.13	0.280	2.240	392

14.12.6 In the system under consideration there will be a different stoichiometric relation for each temperature. For the following temperatures, these relations are expressed basis 100 moles of stack gas carrying 0.30% SO_2.

Eq.	°C	Equation
1	100	$0.30\ SO_2 + 0.60\ H_2S \longrightarrow 0.00072\ S_8 + 0.8942\ S_{Rh} + 0.60\ H_2O$
2	120	$0.30\ SO_2 + 0.60\ H_2S \longrightarrow 0.0042\ S_8 + 0.866\ S_L + 0.60\ H_2O$
3	140	$0.30\ SO_2 + 0.60\ H_2S \longrightarrow 0.0145\ S_8 + 0.784\ S_L + 0.60\ H_2O$
4	160	$0.30\ SO_2 + 0.60\ H_2S \longrightarrow 0.043\ S_8 + 0.556\ S_L + 0.60\ H_2O$
5	180	$0.30\ SO_2 + 0.60\ H_2S \longrightarrow 0.1125\ S_8 + 0.60\ H_2O$
6	200	$0.30\ SO_2 + 0.60\ H_2S \longrightarrow 0.1125\ S_8 + 0.60\ H_2O$

14.12.7 The equilibrium constants could have been based on the above equations but it was decided to base the relations on 1.00 mole SO_2. The advantage desired is the minimum of fractional exponents resulting in easier manual computation. The conversion is obtained by multiplying the above equations by 3.333 to give the following:

Eq.	°C	Equation	°F
1	100	$SO_2 + 2\ H_2S \longrightarrow 0.0024\ S_8 + 2.980\ S_{Rh} + 2\ H_2O$	212
2	120	" $\longrightarrow 0.0140\ S_8 + 2.887\ S_L + 2\ H_2O$	248
3	140	" $\longrightarrow 0.0483\ S_8 + 2.613\ S_L + 2\ H_2O$	284
4	160	" $\longrightarrow 0.143\ S_8 + 1.850\ S_L + 2\ H_2O$	320
5	180	" $\longrightarrow 0.375\ S_8 + 2\ H_2O$	356
6	200	" $\longrightarrow 0.375\ S_8 + 2\ H_2O$	392

14.12.8 The free energy data was obtained from the JANAF tables. The tables are for values of °K at even 100° intervals. The values for the close intervals between 100 and 200°C were interpolated as follows.

14.12.9 Two curves were plotted for values of 300, 400, 500 and 600°K for the following:

(a) Log K for $2H_2S + SO_2 \longrightarrow 3\ S_L + 2\ H_2O$

(b) Log K for S_8

Reading of Log K for the desired centigrade values are interpolated and tabulated for both items. The Log K for S_8 is multiplied by the moles of S_8 and the Log K for the principal reaction is corrected. Tabulated values are shown below.

TABLE XXXI

Eq. No.	°C	Log K Eq.(a)	Log K S_8	Moles S_8	Log K S_8 X Moles S_8	Overall Log K	K
1	100	10.80	-5.10	0.0024	-0.01	10.79	6.16×10^{10}
2	120	9.60	-4.45	0.0140	-0.06	9.54	3.46×10^{9}
3	140	8.85	-3.92	0.0483	-0.17	8.68	4.78×10^{8}
4	160	8.05	-3.50	0.143	-0.50	7.55	3.55×10^{7}
5	180	7.37	-3.08	0.375	-1.15	6.22	1.66×10^{6}
6	200	6.75	-2.73	0.375	-1.03	5.72	5.25×10^{5}

14.12.10 The calculation detail will be shown for the 120°C (248°F) case to illustrate the procedure. Application to the other conditions will be obvious.

$$SO_2 + 2H_2S \longrightarrow 0.0140\ S_8 + 2.887\ S_L + 2H_2O$$

$$K = 3.46 \times 10^9 = \frac{(S_8)^{0.0140}(H_2O)^2}{(SO_2)\ (H_2S)^2}$$

Inlet Gas Composition

SO_2	0.30
H_2S	0.60
H_2O	7.25
Inert	91.85

Basis 100 moles feed gas

x = moles SO_2 reacting

$SO_2 = 0.30-x$

$H_2S = 0.60-2x$

$H_2O = 7.25 + 2x$

$S_8 = 0.0042$ (See vapor pressure table)

Inert = 91.85

Total moles = $100.0042 - x$

$$K = 3.46 \times 10^9 = \frac{(0.0042)^{0.014}(7.25+2x)^2(100.0042-x)^{0.986}}{(0.30-x)(0.60-2x)^2}$$

substituting values of x as follows. Preliminary values leading to the solution are not shown.

X	K
0.293	3.9×10^9
0.292	2.6×10^9

Interpolated solution $x = 0.2925$

Equilibrium gas composition - moles

$SO_2 = 0.30-x = 0.3000-0.2925 = 0.0075$

$H_2S = 0.60-2x = 0.6000-0.5850 = 0.0150$

$H_2O = 7.25+2x = 7.25 + 0.585 = 7.835$

$S_8 = 0.0042 = 0.0042 = 0.0042$

Inert = 91.85 = 91.85 = 91.85

S Liq = 0.900 -0.0075 -0.0150 -8(.0042) = 0.8439

$$\%\ \text{Conversion} = \frac{0.8439 + 8(0.0042)}{0.90} \times 100 = 97.51$$

15.0 APPENDIX

LIST OF EXHIBITS

NOTE: Exhibits material appearing in earlier monthly reports is given with a prefix number designating the number of the report in which it first appeared. Exhibits appearing for the first time in this report are prefixed with the number 11-.

Exhibit No.	Description	Page
4-2	Coke Reduction Unit Heat Rise	110
5-2	Coke Reduction, Process Profile, Case 4	111
5-3	Coke Reduction Flow Sheet, Case 4	112
5-4	Coke Reduction, Process Profile, Case 14	113
5-5	Coke Reduction, Flow Sheet, Case 14	114
5-6	Coke Reduction, Process Profile, Case 15	115
5-7	Coke Reduction, Flow Sheet, Case 15	116
5-8	Coke Reduction, Process Profile, Case 16	117
5-9	Coke Reduction, Flow Sheet, Case 16	118
5-10	Coke Reduction, Summary of Cases	119
5-13	Coke Reduction, Fixed Capital, Case 4	120
5-15	Coke Reduction, Oper. Cost, Case 4	122
5-16	Coke Reduction, Oper. Cost. Case 14	123
5-18	Coke Reduction, Fixed Capital, Case 14	124
5-20	Coke Reduction, Oper. Cost, Case 15	126
5-22	Coke Reduction, Fixed Capital, Case 15	127
5-24a	Coke Reduction, Oper. Cost, Case 16 Factored	128
5-24b	Coke Reduction, Oper. Cost, Case 16 Factored	129
5-25	Coke Reduction, Fixed Capital, Case 16	130
6-1	Equilibrium, $SO_2/CH_4 = 0.80$	131
6-2	Equilibrium, $SO_2/CH_4 = 0.90$	132
6-3	Equilibrium, $SO_2/CH_4 = 1.00$	133
6-4	Equilibrium, $SO_2/CH_4 = 1.10$	134
6-5	Equilibrium, $SO_2/CH_4 = 1.20$	135
6-6	H_2S from CH_4, Efficiency Factor	136
6-7	Equilibrium, $H_2O/CH_4 = 2.0$, $S_2/CH_4 = 2.0$	137
6-8	Equilibrium, $H_2O/CH_4 = 2.0$, $S_2/CH_4 = 2.2$	138
6-9	Equilibrium, $H_2O/CH_4 = 2.2$, $S_2/CH_4 = 2.0$	139

LIST OF EXHIBITS (Cont'd)

Exhibit No.	Description	Page
6-10	Equilibrium, H_2O/CH_4 = 2 2, S_2/CH_4 = 2.2	140
6-11	Equilibrium, H_2O/CH_4 = 2.2, S_2/CH_4 = 2.4	141
6-12	Equilibrium, H_2O/CH_4 = 2.2, S_2/CH_4 = 2.7	142
6-13	Equilibrium, H_2O/CH_4 = 2.2, S_2/CH_4 = 3.0	143
6-14	Equilibrium, H_2O/CH_4 = 2.2, S_2/CH_4 = 3.3	144
6-15	Equilibrium, H_2O/CH_4 = 2.4, S_2/CH_4 = 2.0	145
6-16	Equilibrium, H_2O/CH_4 = 2.4, S_2/CH_4 = 2.2	146
6-17	H_2S Synthesis, CH_4 Efficiency Factor, H_2O/CH_4 = 2.0	147
6-18	H_2S Synthesis, CH_4 Efficiency Factor, H_2O/CH_4 = 2.20	148
6-19	H_2S Synthesis, CH Efficiency Factor, H_2O/CH_4 = 2.40	149
6-20	CH_4 Efficiency at 1300°F	150
6-21A	H_2S Reduction of Gas 2A	151
6-22	Metallic Sulfides Plus SO_2	152
6-23	High Temp. In Situ H_2S Reduction of Gas 2A - Oper. Cost.	153
6-24	H_2S Reduction of Gas 2A, Fixed Capital	154
6-26	H_2S Reduction of Gas 2A, Flow Sheet	155
6-38	Coke Reduction of Gas No. 1, Oper. Cost	156
7-1	Low Temp. Claus - Temp. vs. Catalyst Absorption	157
7-2	Low Temp. Claus - Water vs. Catalyst Absorption	158
7-3	Coke Reduction of Gas 2A, Oper. Cost, Case 14	159
7-4	Flow Sheet - Smelter Gas Cooling	160
7-5	Fixed Capital - Smelter Gas Cooling System	161
8-1	Methane Reduction - Yushkevich - Flow Sheet, Gas No. 1	162
8-2	Methane Reduction - Yushkevich - Process Profile	163
8-3	Methane Reduction - TGS (West) - Flow Sheet, Gas No. 1	164
8-4	Methane Reduction - SO_2/CH_4 Feed Ratio vs. Reaction Temp.	165
8-5	Methane Reduction - Temperature vs. % of SO_2 fed to First Reactor	166
8-6	Methane Reduction - West - Process Profile, Gas No. 1	167

Exhibit No.	Description	Page
8-7	Methane Reduction - Asarco - Flow Sheet, Gas No. 1	168
8-7A	Methane Reduction - Asarco - Process Profile, Gas No. 1	169
8-8	Methane Reduction - West - Fixed Capital, Gas No. 1	170
8-11	Methane Reduction - Kulcsar - Flow Sheet, Gas No. 1	172
8-11A	Methane Reduction - Kulcsar - Process Profile, Gas No. 1	173
8-12	Methane Reduction - Yushkevich - Fixed Capital, Gas No. 1	174
8-14	Methane Reduction - Asarco - Fixed Capital, Gas No. 1	176
8-15	Methane Reduction - Kulcsar - Fixed Capital, Gas No. 1	178
8-20	Methane Reduction - Yushkevich - Oper. Cost, Gas No. 1	180
8-21	Methane Reduction - West, TGS - Oper. Cost, Gas No. 1	181
8-22a	Methane Reduction - Asarco - Oper. Cost, Gas No. 1	182
8-22b	Methane Reduction - Asarco - Factored Oper. Cost, Gas No. 1	183
8-23	Methane Reduction - Kulcsar - Oper. Cost, Gas No. 1	184
8-24	Equilibrium, $SO_2/CH_4 = 2.00$, 10% water	185
8-25	Equilibrium, $SO_2/CH_4 = 2.30$, 10% water	186
8-26	Equilibrium, $SO_2/CH_4 = 2.60$, 10% water	187
8-27	Equilibrium, $SO_2/CH_4 = 3.00$, 10% water	188
8-28	Equilibrium, $SO_2/CH_4 = 3.58$, 10% water	189

LIST OF EXHIBITS

Exhibit No.	Description	Page
8-29	Equilibrium SO_2/CH_4 = 2.00, 0% water	190
8-30	Equilibrium SO_2/CH_4 = 2.60, 0% water	191
8-31	Equilibrium SO_2/CH_4 = 3.10, 0% water	192
9-1	Low Temp. Claus Flow Sheet, Gas No. 2B	193
9-2	Low Temp. Claus Process Profile,Gas No.2B	194
9-3	Low Temp. Claus Draining-Concept Moving Bed	195
9-4	Low Temp. Claus - Catalyst Regenerator	196
9-5	Low Temp. Claus - Catalyst Regen.Flow Sheet	197
9-6	H_2S Synthesis From Reformed Methane-Process Profile	198
9-7	H_2S from Reformed Methane - Flow Sheet	199
9-8	Equilibrium CH_4 Reformer Gas and Sulfur to H_2S	200
9-9	Low Temp. Claus - Staged Converter Bed Flow Sheet	201
9-10	Low Temp. Claus - System Flow Sheet, Case 1D	202
9-11	Low Temp. Claus - Process Profile, Case 1D	203
9-12	Low Temp. Claus - Flow Sheet, Case 2	204
9-13	Low Temp. Claus - Process Profile, Case 2	205
9-14	Low Temp. Claus - Converter Diagram	206
9-15	Low Temp. Claus - Oper. Cost, Case 1	207
9-16	Low Temp. Claus - Fixed Capital, Case 1	208
9-18	Low Temp. Claus - Oper. Cost, Case 1a	210
9-19	Low Temp. Claus - Fixed Capital, Case 1a	211
9-20	Low Temp. Claus - Oper. Cost, Case 1b	212
9-21	Low Temp. Claus - Fixed Capital, Case 1b	213
9-23	Low Temp. Claus - Oper. Cost, Case 1c	215
9-24	Low Temp. Claus - Fixed Capital, Case 1c	216
9-25	Low Temp. Claus - Oper. Cost, Case 1d	218
9-26	Low Temp. Claus - Fixed Capital, Case 1d	219

Exhibit No.	Description	Page
9-27R	Low Temp. Claus - Oper. Cost, Case 2R	221
9-28	Low Temp. Claus - Fixed Capital, Case 2	222
9-28R	Low Temp. Claus - Fixed Capital, Case 2R Revised	224
10-1	CH_4 Reduction - Asarco Flow Sheet, Case 3A	225
10-2	CH_4 Reduction - Asarco - Process Profile, Case 3A	226
10-3	CH_4 Reduction - Process Profile, West, Case 3B	227
10-4	CH_4 Reduction - Flow Sheet, Cases 4A, B & C	228
10-4B	CH_4 Reduction - Flow Sheet, Case 4B	229
10-5	CH_4 Reduction - Process Profile, Asarco, Case 4A	230
10-6	CH_4 Reduction - Process Profile, West, 4B	231
10-7	CH_4 Reduction - Process Profile, Kulcsar, Case 4C	232
10-8	CH_4 Reduction - Summary, Reactor Design Data	233
10-9	Reformed CH_4 Reduction - Flow Sheet, Case 5A	234
10-10	Reformed CH_4 Reduction - Process Profile, Case 5A	235
10-11	Reformed CH_4 Reduction - Flow Sheet, Case 5B	236
10-12	Reformed CH_4 Reduction - Process Profile, Case 5B	237
10-13	Low Temp. Claus - Flow Sheet, Case 6	238
10-14	Low Temp. Claus - Process Profile, Case 6	239
10-15	CH_4 Reduction - Asarco - Oper. Cost, Case 3a	240
10-16	CH_4 Reduction - Asarco Fixed Capital, Case 3a	241
10-17	CH_4 Reduction - West (TGS) - Oper. Cost, Case 3b	243
10-18	CH_4 Reduction - West (TGS) - Fixed Capital, Case 3b	244
10-19	Reformed CH_4 Reduction - Wet Gas - Oper. Cost, Case 5a	246
10-20	Reformed CH_4 Reduction - Wet Gas - Fixed Capital Case 5a	247
10-21	Reformed CH_4 Reduction - Dry Gas - Oper. Cost, Case 5b	249
10-22	Reformed CH_4 Reduction - Dry Gas - Fixed Capital, Case 5b	250
10-23	CH_4 Reduction, Asarco - Oper. Cost, Case 4a	252
10-24	CH_4 Reduction, Asarco - Fixed Capital, Case 4a	253
10-25	CH_4 Reduction, West (TSG) - Oper. Cost, Case 4b	255
10-26	CH_4 Reduction, West (TSG) - Fixed Capital, Case 4b	256
10-27	CH_4 Reduction, Kulcsar - Oper. Cost, Case 4c	258
10-28	CH_4 Reduction, Kulcsar - Fixed Capital, Case 4c	259

LIST OF EXHIBITS (Cont'd)

Exhibit No.	Description	Page
10-29a	Low Temp. Claus - Oper. Cost, Case No. 6	261
10-29b	Low Temp. Claus - Oper. Cost, Case No. 6, Factored	262
10-30	Low Temp. Claus - Fixed Capital, Case No. 6	263
10-31b	Low Temp. Claus - Oper. Cost, Case 2R, Factored	265
11-1	Cost of Natural Gas - Survey of U. S.	266
11-2	Equilibrium of Some Reactions in the SO_2/CH_4 System	268
11-3	Equilibrium $SO_2/CH_4 = 1.33$, $H_2O = 0$	269
11-4	Equilibrium $SO_2/CH_4 = 1.33$, $H_2O = 0.53$.	270
11-5a	Comparison Kulesar Data with Equilibria	271
11-5b	Averbukh's Data on CH_4 Reduction	272
11-6	Low Temp. Claus - Switch Reactors - Oper. Cost, Case 2B	273
11-7	Low Temp. Claus - Switch Reactors - Fixed Capital, Case 2B	274
11-8	Low Temp. Claus - Switch Reactors - Flow Sheet, Case 2B	276
11-9	Low Temp. Claus - Switch Reactors - Process Profile, Case 2B	277
11-10	Low Temp. Claus - Switch Reactor - Detail Sketch	278
11-11	Low Temp. Claus - Moving Bed Reactor - Oper. Cost, Case 2A	279
11-12	Low Temp. Claus - Moving Bed Reactor - Fixed Capital, Case 2A.	280
11-13	Low Temp. Claus = Cross Flow Reactor Sketch, Case 2A	282

COKE REDUCTION UNIT HEAT RISE - Adiabatic

Exhibit 4-2

90% SO_2 reduction
$H_2S + CO + H_2$ equivalent to unreduced SO_2
Feed gas in at 950° C Case 1-15
Feed gas in at 227° C Case 16

CASE	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16
Feed gas Comp. %																
SO_2	100.00	90.00	90.00	83.33	83.33	90.00	72.00	60.00	91.00	79.00	2.90	15.80	15.32	14.58	9.58	4.50
H_2O		10.00	10.00	16.67	16.67		8.00	6.66		14.00	0.10		3.06	2.92	28.14	
O_2						10.00				7.00	14.30	1.20	1.16	2.31	2.74	16.50
Air							20.00	33.34								
CO											0.60					
CO_2											1.70					
CH_4									9.00							
N_2											80.40	83.00	80.46	80.19	59.27	79.00
Product gas Comp. %																
SO_2	6.46	6.23	6.02	6.06	5.71	6.02	5.37	4.68	6.46	5.79	0.29	1.45	1.43	1.37	0.98	
H_2S		6.92		12.12			5.90	5.14	12.77	10.26	0.10		2.87	2.74	1.96	
CO	12.90	5.54	5.35			12.04	4.84	4.21			0.48	2.91				24.09
H_2			6.69		11.42					1.32						
CO_2	51.61	56.74	54.85	61.61	57.16	54.85	51.33	47.44	58.10	61.71	18.67	12.73	15.42	15.85	12.57	6.02
S_2	29.03	24.57	27.09	21.21	25.71	27.09	20.91	18.22	22.67	20.92	1.25	6.54	5.01	4.79	3.46	1.94
N_2							11.65	20.31			80.21	76.37	75.27	75.25	59.27	67.95
H_2O															26.18	
Outlet gas temp. °C	913	1060	955	1240	990	1413	1290	1465	1255	1519	2255	1046	1137	1244	1257	1283
Reactor temp. rise °C	-37	+110	+5	+290	+40	+480	+340	+515	+305	+569	+1305	+96	+187	+295	+307	956
Lb. carbon/lb. sulfur	0.375	0.375	0.375	0.375	0.375	0.417	0.397	0.419	0.300	0.408	2.185	0.403	0.403	0.434	0.479	1.750
Rel. gas vol. Case 1 = 1.00	1.00	1.11	1.11	1.20	1.20	1.11	1.39	1.67	1.10	1.27	34.5	6.32	6.53	6.86	10.15	22.22
Ht. reaction cal/mol SO_2	-3107	-6820	-4350	-11040	-5294	-14720	-12870	-18980	-10070	-18290	-524300	-9940	-15960	-23690	-31680	-224300

EXH. 5-2

Coke Reduction 500 N.T. Sulfur / Day

Case 4 Process Profile

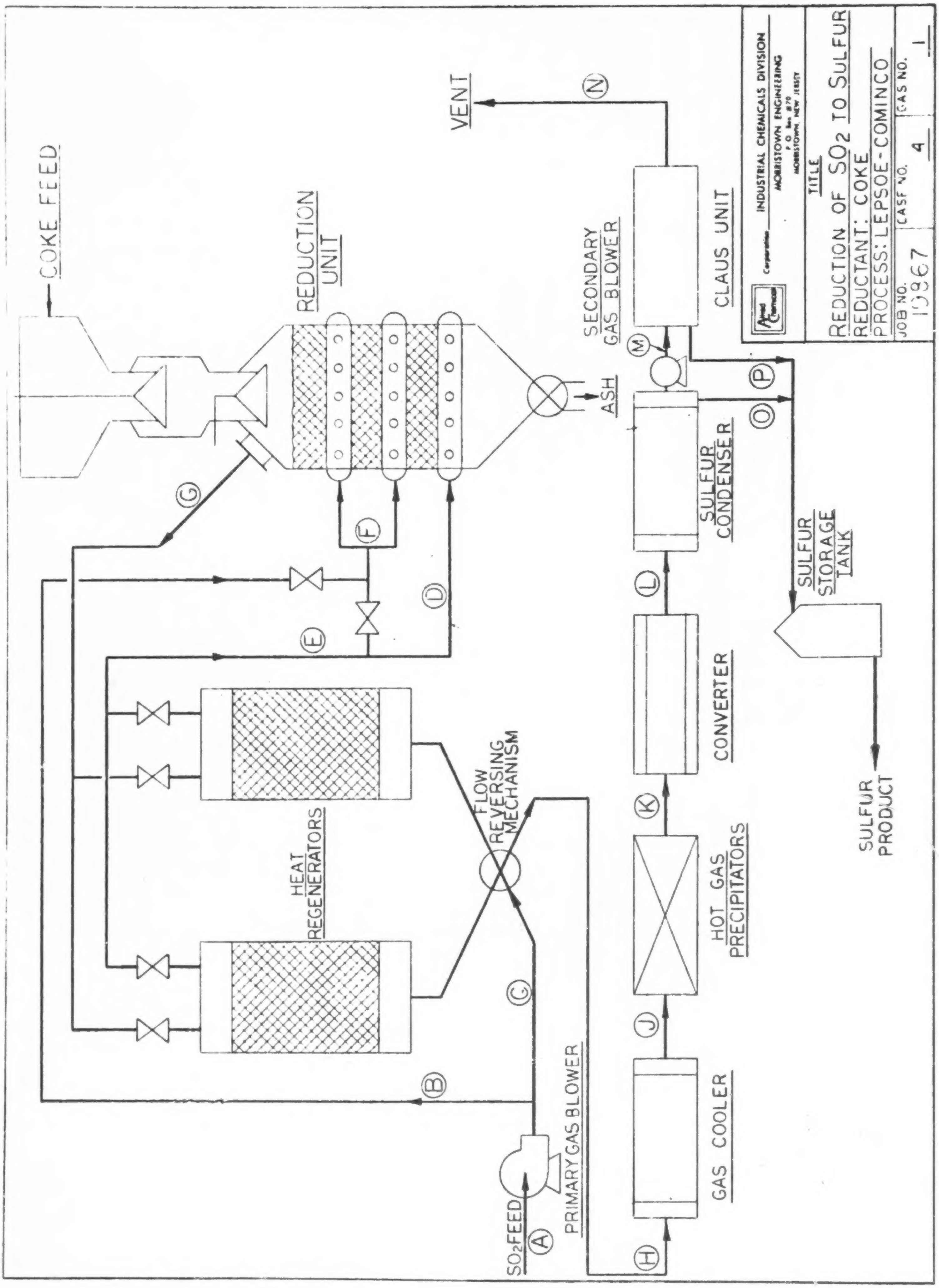
Process Points	A	B	C	D	E	F	G	H	J	K	L	M	N
Temp °C	20	20	20	1100	1100	842	1187	500	450	425	400	220	150
Vol. S.C.F.M.	9450	1100	8350	4170	4170	5270	12990	12990	12990	12990	13130	10020	9480
Vol. cfm @ T°	10140	1180	8960	21000	21000	21520	70220	39580	34770	33570	32370	18090	14690
% SO_2	83.30	83.30	83.30	83.30	83.30	83.30	6.04	6.04	6.04	6.04	3.81	5.00	0.16
% H_2O	16.70	16.70	16.70	16.70	16.70	16.70					4.37	5.72	16.27
% S as S_2								21.50	21.50	21.50	24.26	0.75	0.20
% CO_2								60.39	60.39	60.39	59.95	78.56	83.05
% H_2S								12.07	12.07	12.07	7.61	9.97	0.32
% N_2													

Coke Feed	9.30 N.T/Hr.	85.0 % Carbon
Ash	1.40 N.T/Hr.	5.0 % Carbon

Sulfur Distribution N.T. / Day

POINT Ⓞ	Sulfur Cooler	399.0
POINT Ⓟ	Claus Unit	101.0
	Vent	5.3
	% S Recovery	98.95
	% Heat Loss	7.0

EXHIBIT 5-3


COKE FEED
REDUCTION UNIT
ASH
VENT
Ⓝ
CLAUS UNIT
SECONDARY GAS BLOWER
Ⓜ
SULFUR CONDENSER
Ⓞ
Ⓟ
SULFUR STORAGE TANK
SULFUR PRODUCT
Ⓛ
CONVERTER
Ⓚ
HOT GAS PRECIPITATORS
Ⓙ
GAS COOLER
Ⓗ
Ⓖ
Ⓕ
Ⓔ
Ⓓ
HEAT REGENERATORS
FLOW REVERSING MECHANISM
Ⓑ
PRIMARY GAS BLOWER
SO_2 FEED
Ⓐ

Allied Chemical Corporation
INDUSTRIAL CHEMICALS DIVISION
MORRISTOWN ENGINEERING
P. O. Box 870
MORRISTOWN, NEW JERSEY

TITLE
REDUCTION OF SO_2 TO SULFUR
REDUCTANT: COKE
PROCESS: LEPSOE-COMINCO

JOB NO.	CASF NO.	CAS NO.
19967	4	1

DRAWING NO. 37177

EXHIBIT 5-4

Coke Reduction 200 N.T. Sulfur / Day EXH. 5-4

Process Profile Case 14 2A Conv. Gas + Air + H_2O - Indirect Cooling

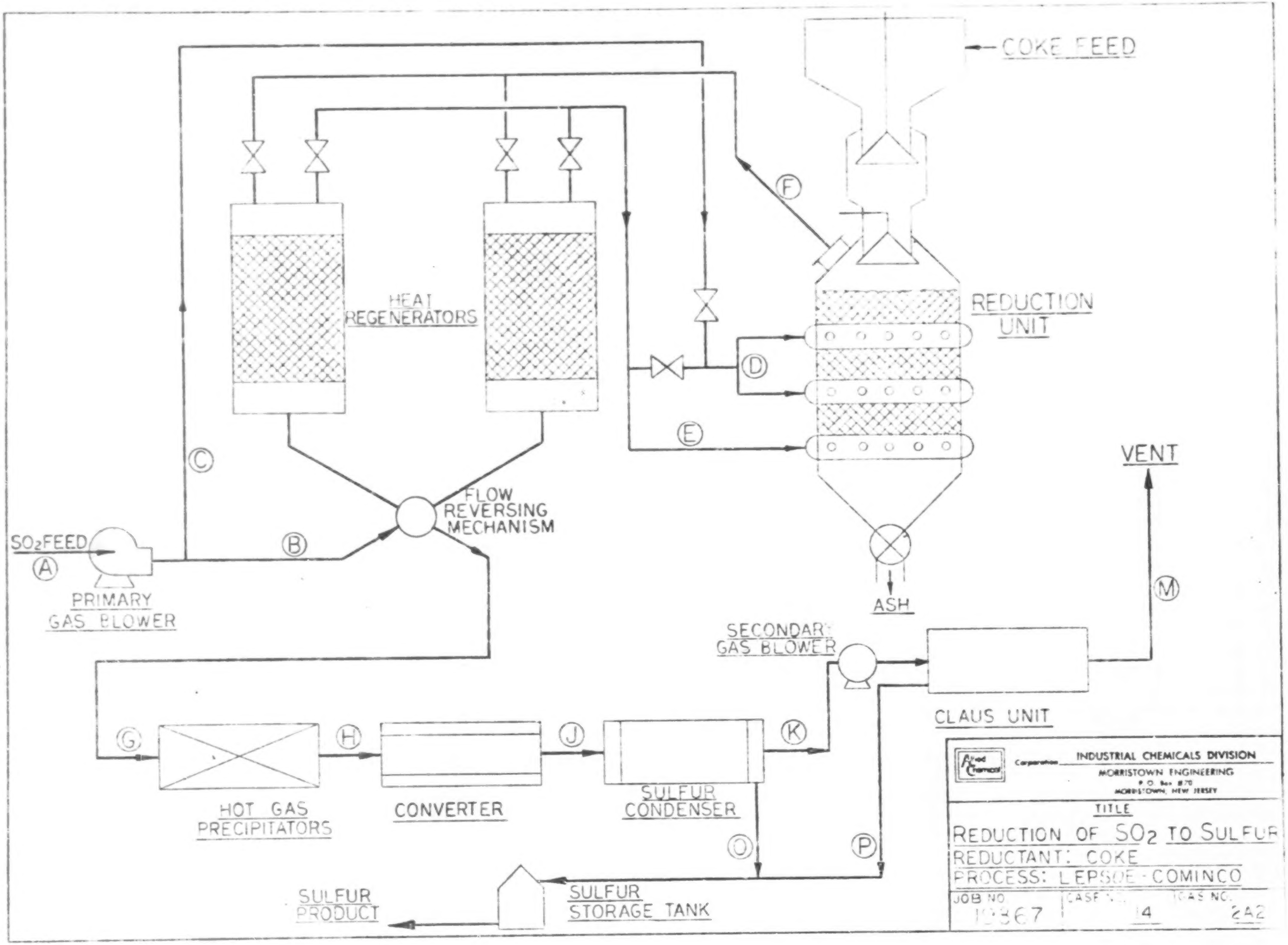
Process Points	A	B	C	D	E	F	G	H	J	K	L	M	N
Temp °C	160	160	160	932	1100	1196	450	425	420	220		150	
Vol. SCFM.	22240	20060	2180	12210	10030	23720	23720	23720	23780	22590		22290	
Vol. cfm @ T°	35270	31780	3460	53900	50350	127640	62800	60640	60040	40800		34530	
% SO_2	14.58	14.58	14.58	14.58	14.58	1.37	1.37	1.37	0.73	0.77		0.06	
% H_2O	2.92	2.92	2.92	2.92	2.92				1.27	1.34		2.77	
% S as S_2						4.79	4.79	4.79	5.73	0.79		0.20	
% CO_2						15.85	15.85	15.85	15.80	16.63		16.84	
% H_2S						2.74	2.74	2.74	1.46	1.54		0.13	
% N_2	80.19	80.19	80.19	80.19	80.19	75.25	75.25	75.25	75.01	78.93		80.00	
% O_2	2.31	2.31	2.31	2.31	2.31								

Coke Feed	4.30 N.T./Hr.	85.0 % Carbon	
Ash	0.65 N.T./Hr.	5.0 % Carbon	

	Sulfur Distribution	N.T. / Day
Point Ⓞ	Sulfur Cooler	152.0
Point Ⓟ	Claus Unit	48.0
	Vent	8.2

%S Recovery 96.08

% Heat Loss 2.0


COKE FEED
HEAT REGENERATORS
REDUCTION UNIT
FLOW REVERSING MECHANISM
SO₂ FEED
A
B
C
D
E
F
G
H
J
K
M
O
P
PRIMARY GAS BLOWER
ASH
VENT
SECONDARY GAS BLOWER
CLAUS UNIT
HOT GAS PRECIPITATORS
CONVERTER
SULFUR CONDENSER
SULFUR PRODUCT
SULFUR STORAGE TANK

Allied Chemical Corporation INDUSTRIAL CHEMICALS DIVISION
MORRISTOWN ENGINEERING
P.O. Box 870
MORRISTOWN, NEW JERSEY

TITLE
REDUCTION OF SO_2 TO SULFUR
REDUCTANT: COKE
PROCESS: LEPSOE - COMINCO

JOB NO.	CASE NO.	CAS NO.
10367	14	2A2

DRAWING NO. 37178

Coke Reduction 200 N.T. Sulfur / Day EXH. 5-6

Case 15 Process Profile 2A Conv. Gas + Air - Water Injection Cooling to 900°F

Process Points	A	B	C	D	E	F	G	H	J	K		M	N
Temp °C	160	160	160	929	1100	1207	450	425	425	220		150	
Vol. SCFM	35500	31950	3550	19530	15970	37080	37080	37080	37080	36140		35650	
Vol. cfm @ T°	56310	50680	5630	86000	80310	201000	98280	94800	94800	65280		55240	
% SO_2	9.85	9.85	9.85	9.85	9.85	0.94	0.94	0.94	0.94	0.96		0.22	
% H_2O	28.14	28.14	28.14	28.14	28.14	25.07	25.07	25.07	25.07	25.72		27.57	
% S as S_2						3.30	3.30	3.30	3.30	0.79		0.20	
% CO_2						12.06	12.06	12.06	12.06	12.37		12.54	
% H_2S						1.88	1.88	1.88	1.88	1.93		0.45	
% N_2	59.27	59.27	59.27	59.27	59.27	56.75	56.75	56.75	56.75	58.23		59.02	
% O_2	2.74	2.74	2.74	2.74	2.74								

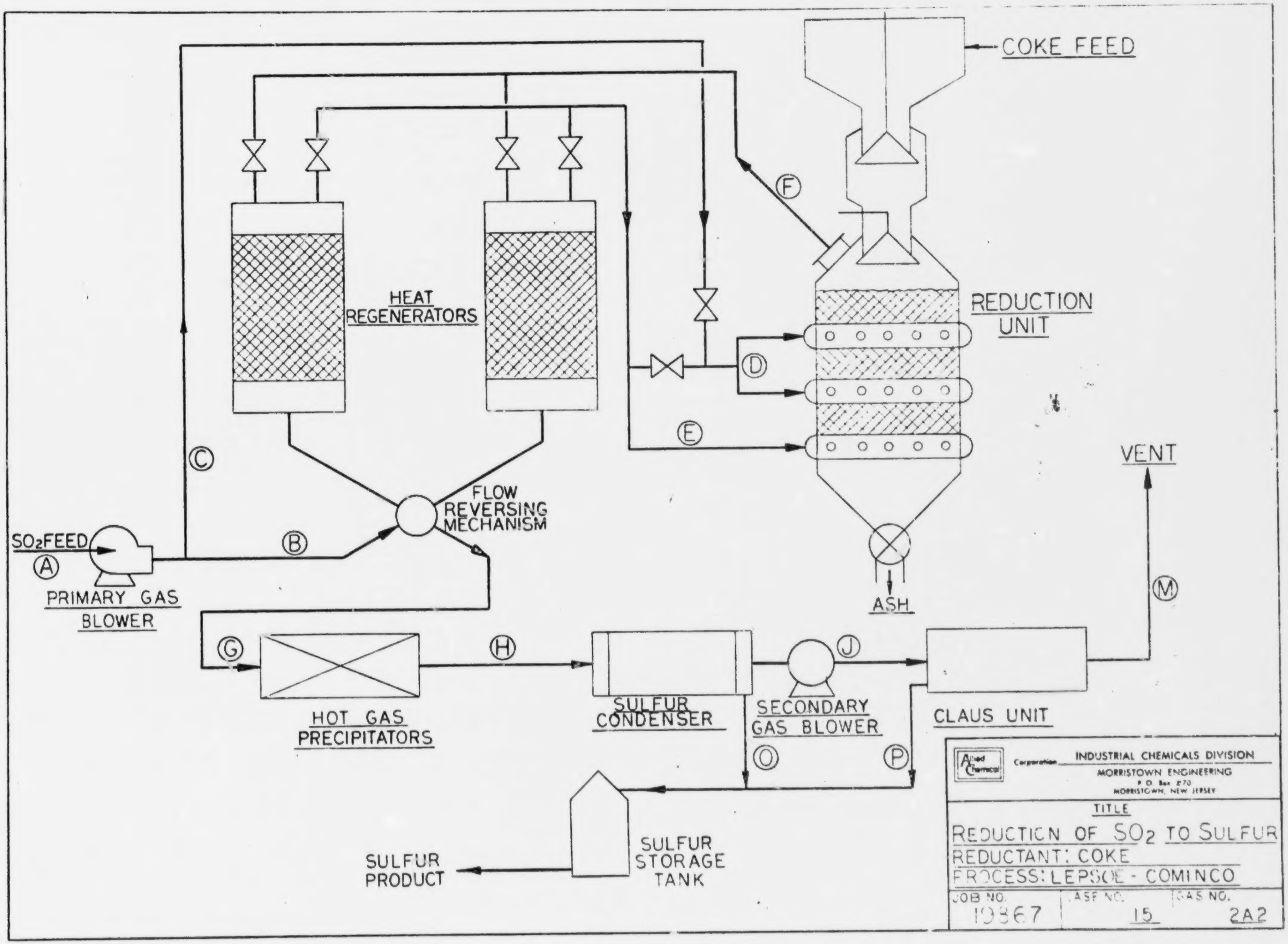
Coke Feed 4.75 N.T./Hr. 85.0 % Carbon

Ash 0.72 N.T./Hr. 50 % Carbon

Sulfur Distribution	N.T. / Day
Point Ⓞ Sulfur Cooler	121.0
Point Ⓟ Claus Unit	79.0
Vent	24.4
% Sulfur Recovery	89.1
% Heat Loss	7.0

EXHIBIT 5-6

EXHIBIT 5-7


COKE FEED
HEAT REGENERATORS
REDUCTION UNIT
FLOW REVERSING MECHANISM
SO_2 FEED
Ⓐ
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Ⓞ
Ⓟ
PRIMARY GAS BLOWER
ASH
VENT
HOT GAS PRECIPITATORS
SULFUR CONDENSER
SECONDARY GAS BLOWER
CLAUS UNIT
SULFUR STORAGE TANK
SULFUR PRODUCT

Allied Chemical Corporation INDUSTRIAL CHEMICALS DIVISION
MORRISTOWN ENGINEERING
P.O. Box 270
MORRISTOWN, NEW JERSEY

TITLE

REDUCTION OF SO_2 TO SULFUR
REDUCTANT: COKE
PROCESS: LEPSOE - COMINCO

JOB NO.	CASE NO.	GAS NO.
10367	15	2A2

DRAWING NO. 37179

Coke Reduction 500 N.T. Sulfur / Day EXH. 5-8

Process Profile Case 16 2B Smelter Gas

Process Points	A	B	C	D	E	F	G	H	J	K	L	M	N
Temp °C	227	227	227	1195	860	1745	475	450	556	150			
Vol. SCFM.	185500	111300	74200	129400	203600	191300	191300	191300	189700	185800			
Vol. cfm @ T°	339800	203900	135900	695800	844900	1414100	524200	506600	576100	287800			
% SO_2	4.50	4.50	4.50	0.00	1.64	1.75	1.75	1.75	0.06	0.06			
% O_2	16.50	16.50	16.50	0.00	6.01								
% S as S_2				1.94	1.23	1.31	1.31	1.31	2.17	0.12			
% CO_2				6.02	3.83	16.87	16.87	16.87	20.42	20.84			
% CO				24.09	15.31	3.49	3.49	3.49	0.12	0.12			
% N_2	7900	79.00	79.00	67.95	71.98	76.58	76.58	76.58	77.23	78.86			

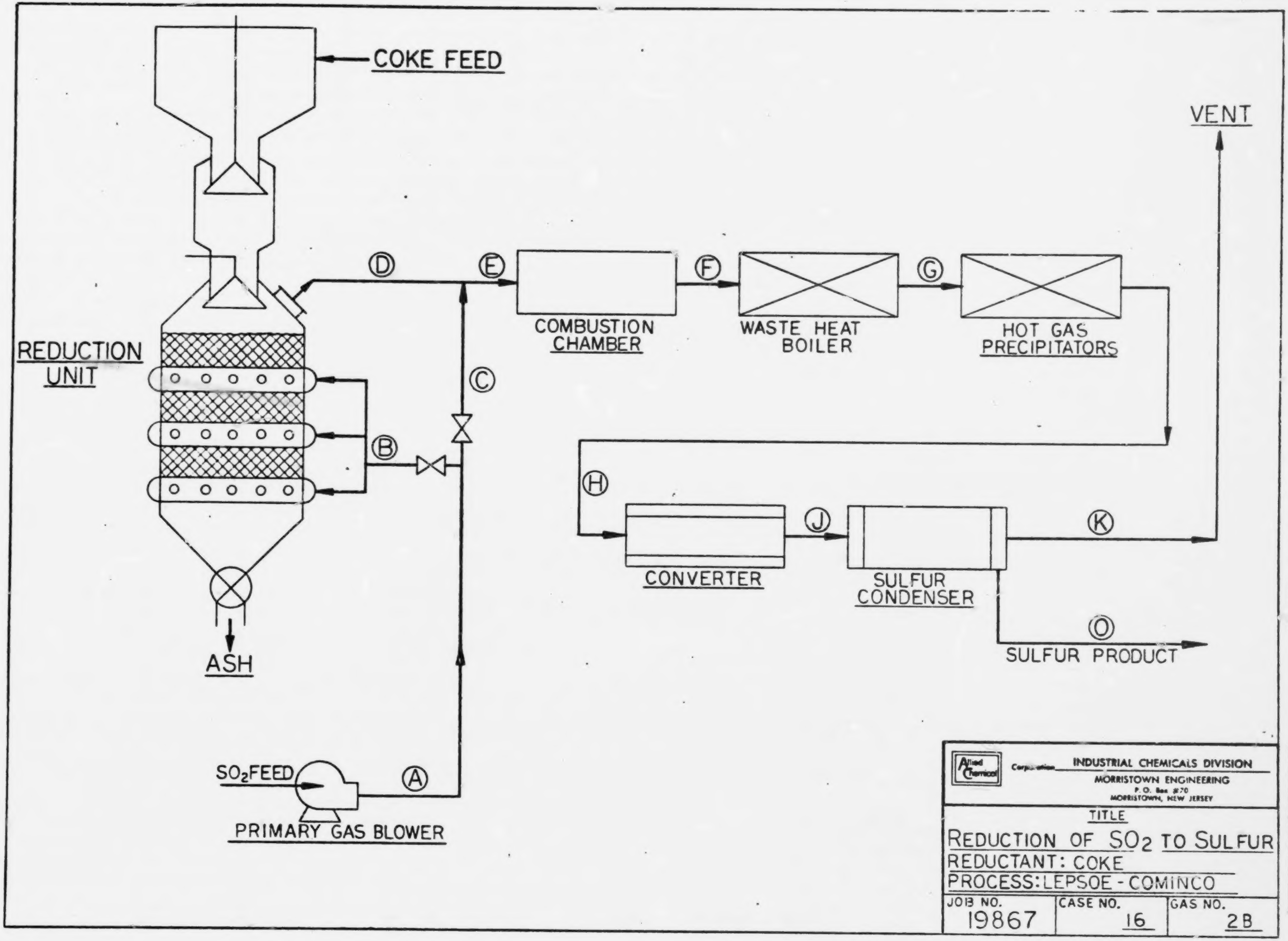
Coke Feed	43.4	N.T./Hr.	85.0 %	Carbon
Ash	6.5	N.T./Hr.	50 %	Carbon

Sulfur Distribution	N.T. / Day
[illegible] D) Sulfur Cooler	500
Claus Unit	—
Vent	36

% Recovery - 93.3

EXHIBIT 5-8

EXHIBIT 5-9


COKE FEED
VENT
Ⓓ
Ⓔ
Ⓕ
Ⓖ
COMBUSTION CHAMBER
WASTE HEAT BOILER
HOT GAS PRECIPITATORS
REDUCTION UNIT
Ⓒ
Ⓑ
Ⓗ
CONVERTER
Ⓙ
SULFUR CONDENSER
Ⓚ
Ⓞ
SULFUR PRODUCT
ASH
SO_2FEED
Ⓐ
PRIMARY GAS BLOWER
Allied Chemical

Corporation INDUSTRIAL CHEMICALS DIVISION
MORRISTOWN ENGINEERING
P. O. Box #70
MORRISTOWN, NEW JERSEY

TITLE
REDUCTION OF SO_2 TO SULFUR
REDUCTANT: COKE
PROCESS: LEPSOE - COMINCO

JOB NO.	CASE NO.	GAS NO.
19867	16	2B

DRAWING NO. 3/180

COKE REDUCTION OF SO_2 — EXH 5-10

PLANT PROJECTION SUMMARY

CASE No.		4	14	15	16
CAPACITY - TONS SULFUR /DAY		500	200	200	500
COKE FEED - TONS /Hr.		9.3	4.3	4.75	43.4
ASH - TONS/Hr.		1.4	0.65	0.72	6.5
% SULFUR RECOVERY		98.95	96.08	89.10	93.30
REDUCTION UNIT	No. UNITS	1	2	3	10
	DIA. - Ft.	13.9	13.3	13.5	13.9
	BED DEPTH - Ft.	12.0	12.0	12.0	12.0
REGENERATOR	No. PAIRS	1	1	1	NONE
	DIA. - Ft.	9.1	12.3	15.4	
	BED DEPTH - Ft.	10.0	10.0	10.0	
GAS COOLER	No. UNITS	1	NONE	NONE	NONE
PRECIPITATOR	No. UNITS	1	1	1	1
CATALYST BED	No. UNITS	1	1	NONE	8
	DIA. - Ft.	14.8	20.6		30.0
	BED DEPTH - Ft.	6.0	6.0		6.0
SULFUR CONDENSER	No. UNITS	1	1	1	1
CLAUS UNIT	No. BEDS	2	2	4	NONE
	DIA - Ft.	14.8	20.6	18.2	
	BED DEPTH - Ft.	6.0	6.0	6.0	
SO_2 FEED	% SO_2	83.30	14.58	9.85	4.50
	SCFM	9450	22 240	35500	185500

11/8/68

EXHIBIT 5-13

CASE 4 (GAS NO. 1)

SUMMATION OF MAJOR EQUIPMENT

1.	Primary gas blower - multi-stage centrifugal blower to handle 9500 SCFM and 8 psig pressure boost, complete with 600 HP motor.	$ 30,000
2.	Flow reversing mechanism - vertical four-way piston valve with inlet and outlet vestibules complete with hydraulic drive operating mechanism. This is a special design developed by Allied Chemical.	$ 30,000
3.	Heat regenerators - two high duty refractory lined steel vessels 9'-3" inside diameter packed with 10 feet of 1 1/2" dia. refractory balls as a heat transfer medium, complete with packing support grate of meehanite and gas distribution baffles.	$190,000
4.	Reduction unit - super duty refractory lined steel vessel 14' inside diameter with a bed depth of 12' complete with coke feed mechanism rated at 10 NT/hr., coke feed bin, double gate for pressure balance and ash discharge system.	$200,000
5.	Hot gas valves - four - 2' diameter and 2 - 1 foot diameter water cooled butterfly valves with refractory lined discs. These are specially designed valves developed by Allied Chemical and are included as major equipment.	$ 35,000
6.	Gas cooler - shell and tube gas to air heat exchangers with steel and alonized tubes will reduce gas temperature 50°C to stay within temperature limits of precipitator.	$ 30,000
7.	Hot gas precipitators - plate type precipitators of carbon steel construction with dust removal drag and screw conveyor. Two units in parallel each rated at 4800 SCFM and 450°C.	$150,000
8.	Converter - refractory lined steel vessel 14'-0" I.D. packed to a depth of 6' with alumina type catalyst.	$ 50,000
9.	Sulfur condenser - horizontal type waste heat boiler with integral steam drum to produce 50 psig steam. Unit will be installed with a 4° incline for sulfur drainage.	$ 50,000
10.	Secondary gas blower - centrifugal gas blower to handle 10,000 SCFM at 220°C inlet temperature complete with 500 HP motor.	$ 30,000

11.	Partial Claus sulfur recovery unit - consisting of a two stage converter, two sulfur condensers, coalescer, incinerator and exit stack. Unit is designed for a throughput of 10,000 SCFM with high temperature incineration for removal of residual H_2S and 150' exit stack in keeping with Allied Chemical's standard design.-------------------------------	$137,000
12.	Sulfur storage tank - vertical steel tank 5,000 ton capacity with steam coils and external insulation.-----------------	$ 70,000
	Total major purchased equipment---------------------------	$1,002,000
	Total Fixed Capital--------------------------------------	$3,000,000

EXHIBIT 5-15
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: Coke CASE NO. 4

PROCESS: Lepsoe - Cominco at Trail GAS NO. 1

Fixed Capital, Battery Limits Basis $ 1,750,000

Sulfur Production 200 NTD x 330 Days = 66,000 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	MCF	$.30 Avg.	$	$
	Coke, 3" Furnace Grade	0.45 NT	$24. Avg.	10.85	715.0
1.	Total Raw Materials	-	-	$ 10.85	$ 715.0
2.	Direct Labor	18,000 MH	$3.50	.96	63.0
3.	Supervision	1.	$10,000	.15	10.0
4.	Maintenance	5% FC	-	1.32	87.5
5.	Plant Supplies	10%(L2+L3)	-	.11	7.3
	Utilities:				
6. a)	Electricity	80. KWH	$.01	.80	52.8
b)	Cooling Water 80 gpm	0.58 MG	.05	.03	2.0
c)	Boiler Feed Water	- MG	.25	-	-
d)	Fuel Gas	0.9 MCF	.30	.27	17.4
7.	Total Utilities	-	-	1.10	72.2
8.	Other Direct (Catalyst)	5 NTY	$100.	.01	0.5
9.	TOTAL DIRECT	-	-	$ 14.50	$ 955.9
10.	Payroll Burden	30% L2	-	.29	18.9
11.	Plant Overhead	70%(L2+L3)	-	.77	51.1
12.	Pack & Ship	1000 MH	3.90	.06	3.9
13.	Waste Disposal	1000 MH	3.90	.06	3.9
14.	Other Indirect	-	-	-	-
15.	TOTAL INDIRECT	-	-	$ 1.18	$ 77.8
16.	Depreciation	10% FC	-	2.65	175.0
17.	Taxes	2% FC	-	.53	35.0
18.	Insurance	1% FC	-	.26	17.5
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$ 3.44	$ 227.5
21.	TOTAL OPERATING	-	-	$ 19.12	$1,261.2
22.	Credits HP Steam	.96 M Lbs.	.50	.48	31.6
23.	NET OPERATING	-	-	$ 18.64	$1,229.6

EXHIBIT 5-16
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: Coke — CASE NO. 14
PROCESS: Lepsoe - No heat recovery — GAS NO. 2A

Fixed Capital, Battery Limits Basis $ 4,200,000
Sulfur Production 195 NTD x 330 Days = 64,400 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	MCF	$.30 Avg.	$ -	$ -
	Coke	0.516 NT	24. Avg.	12.40	800.0
1.	Total Raw Materials	-	-	$ 12.40	$ 800.0
2.	Direct Labor	18,000 MH	$3.50	.98	63.0
3.	Supervision	-	$10,000	.15	10.0
4.	Maintenance	5% FC	-	3.26	210.0
5.	Plant Supplies	10%(L2+L3)	-	.11	7.3
	Utilities:				
6. a)	Electricity 3900 KW	470 KWH	$.01	4.70	302.0
b)	Cooling Water 300 gpm	2.0 MG	.05	.10	6.4
c)	Boiler Feed Water	- MG	.25	-	-
d)	Fuel Gas	4.3 MCF	.30	1.29	82.0
7.	Total Utilities	-	-	6.09	390.4
8.	Other Direct (Catalyst)	5 NTY	$100	.01	0.5
9.	TOTAL DIRECT	-	-	$ 23.00	$1,481.2
10.	Payroll Burden	30% L2	-	.29	18.9
11.	Plant Overhead	70%(L2+L3)	-	.79	51.0
12.	Pack & Ship	1000 MH	3.90	.06	3.9
13.	Waste Disposal	1000 MH	3.90	.06	3.9
14.	Other Indirect	-	-	-	-
15.	TOTAL INDIRECT	-	-	$ 1.20	$ 77.7
16.	Depreciation	10% FC	-	6.53	420.0
17.	Taxes	2% FC	-	1.31	84.0
18.	Insurance	1% FC	-	.65	42.0
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$ 8.49	$ 546.0
21.	TOTAL OPERATING	-	-	$ 32.69	$2,104.9
22.	Credits HP Steam	1.M Lb.	.50	.50	32.2
23.	NET OPERATING	-	-	$ 32.19	$2,072.7

EXHIBIT 5-18

CASE 14 (GAS NO. 2A)

SUMMATION OF MAJOR EQUIPMENT

1.	Primary gas blower - multi-stage centrifugal blower to handle 22,500 SCFM at 160°C with a Δp of 8 psi.	$ 60,000
2.	Flow reversing mechanism - vertical four-way piston valve with inlet and outlet vestibules complete with hydraulic drive operating mechanism. This is a special design developed by Allied Chemical.	$ 50,000
3.	Heat regenerators - two high duty refractory lined steel vessels 12'-4" inside diameter packed with 10 feet of 1 1/2" dia. refractory balls as a heat transfer medium, complete with packing support grate of meehanite and gas distribution baffles.	$230,000
4.	Reduction unit-2 super duty refractory lined steel vessels 14' inside diameter with a bed depth of 13'4" complete with coke feed mechanism rated at 2.2 NT/hr., coke feed bin, double gate for pressure balance and ash discharge system.	$400,000
5.	Hot gas valves - four - 3' diameter and 2 - 18" diameter water cooled butterfly valves with refractory lined discs. These are specially designed valves developed by Allied Chemical and are included as major equipment.	$ 45,000
6.	Hot gas precipitators - plate type precipitators of carbon steel construction with dust removal drag and screw conveyor. Two units in parallel each rated at 11,250 SCFM and 450°C.	$220,000
7.	Converter - refractory lined steel vessel 20'-6" I.D. packed to a depth of 6' with alumina type catalyst.	$ 65,000
8.	Sulfur condenser - horizontal type waste heat boiler with integral steam drum to produce 50 psig steam. Unit will be installed with a 4° incline for sulfur drainage.	$ 73,000
9.	Secondary gas blower - centrifugal gas blower to handle 23,000 SCFM at 220°C inlet temperature complete with 1000 HP motor.	$ 40,000
10.	Partial Claus sulfur recovery unit - consisting of a two stage converter, two sulfur condensers, coalescer, incinerator and exit stack. Unit is designed for a throughput of 23,000 SCFM with high temperature incineration for removal of residual H_2S and 150' exit stack in keeping with Allied Chemical's standard design.	$180,000

EXH. 5-18 (Cont.)

11.	Sulfur storage tank - vertical steel tank 2,000 ton capacity with steam coils and external insulation.	$ 40,000
	Total major purchased equipment	$1,403,000
	Total Fixed Capital	$4,200,000

EXHIBIT 5-20
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: Coke CASE NO. 15
PROCESS: Lepsoe, Hot Feed Quenched GAS NO. 2A2

Fixed Capital, Battery Limits Basis $ 5,200,000
Sulfur Production 180 NTD x 330 Days = 59,500 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	- MCF	$.30 Avg.	$ -	$ -
	Coke 3" Furnace Grade	0.57	$24. Avg.	13.70	815.0
1.	Total Raw Materials	-	-	$13.70	$ 815.0
2.	Direct Labor	18,000 MH	$3.50	1.06	63.0
3.	Supervision	1.	$10,000	.17	10.0
4.	Maintenance	5% FC	-	4.37	260.0
5.	Plant Supplies	10%(L2+L3)	-	.12	7.3
	Utilities:				
6. a)	Electricity 6200 KW	825 KWH	$.01	8.25	490.1
b)	Cooling Water 400 gpm	3.2 MG	.05	.16	9.5
c)	Boiler Feed Water	- MG	.25	-	-
d)	Fuel Gas	8.0 MCF	.30	2.40	143.0
7.	Total Utilities	-	-	10.81	642.6
8.	Other Direct (Catalyst)	5 NTY	$100	.01	0.5
9.	TOTAL DIRECT	-	-	$30.24	$1,798.4
10.	Payroll Burden	30% L2	-	.32	18.9
11.	Plant Overhead	70%(L2+L3)	-	.86	51.1
12.	Pack & Ship	1000 MH	3.90	.06	3.9
13.	Waste Disposal	1000 MH	3.90	.07	3.9
14.	Other Indirect	-	-	-	-
15.	TOTAL INDIRECT	-	-	$ 1.31	$ 77.8
16.	Depreciation	10% FC	-	8.75	520.0
17.	Taxes	2% FC	-	1.75	104.0
18.	Insurance	1% FC	-	.88	52.0
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$11.38	$ 676.0
21.	TOTAL OPERATING	-	-	$42.93	$2,552.2
22.	Credits LP Steam	1.0 M Lb.	.25	.25	14.9
23.	NET OPERATING	-	-	$42.68	$2,537.3

EXHIBIT 5-22

CASE 15

SUMMATION OF MAJOR EQUIPMENT

1.	Primary gas blower - multi-stage centrifugal blower to handle 35,500 SCFM at 160°C with a Δp of 8 psi.	$ 80,000
2.	Flow reversing mechanism - vertical four-way piston valve with inlet and outlet vestibules complete with hydraulic drive operating mechanism. This is a special design developed by Allied Chemical.	$ 60,000
3.	Heat regenerators - two high duty refractory lined steel vessels 15'-6" inside diameter packed with 10 feet of 1 1/2" dia. refractory balls as a heat transfer medium; complete with packing support grate of meehanite and gas distribution baffles.	$270,000
4.	Reduction unit - 2 super duty refractory lined steel vessels 14' inside diameter with a bed depth of 13'-6" complete with coke feed mechanism rated at 1.6 NT/hr., coke feed bin, double gate for pressure balance and ash discharge system.	$600,000
5.	Hot gas valves - four - 3'-6" diameter and 3 - 2' diameter water cooled butterfly valves with refractory lined discs. These are specially designed valves developed by Allied Chemical and are included as major equipment.	$ 60,000
6.	Hot gas precipitators - plate type precipitators of carbon steel construction with dust removal drag and screw conveyor. Two units in parallel each rated at 18,000 SCFM and 450°C.	$260,000
7.	Sulfur condenser - horizontal type waste heat boiler with integral steam drum to produce 50 psig steam. Unit will be installed with a 4° incline for sulfur drainage.	$ 97,000
8.	Secondary gas blower - centrifugal gas blower to handle 35,500 SCFM at 220°C inlet temperature complete with 1500 HP motor.	$ 50,000
9.	Partial Claus sulfur recovery unit - consisting of a two stage converter, two sulfur condensers, coalescer, incinerator and exit stack. Unit is designed for a throughput of 35,500 SCFM with high temperature incineration for removal of residual H_2S and 150' exit stack in keeping with Allied Chemical's standard design.	$220,000
10.	Sulfur storage tank - vertical steel tank 2,000 ton capacity with steam coils and external insulation.	$ 40,000
	Total major purchased equipment	$1,737,000

EXHIBIT 5-24a
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: Coke CASE NO. 16 (Factored)
PROCESS: Lepsoe - Cominco GAS NO. 2B

Fixed Capital, Battery Limits Basis $ 20,500,000
Sulfur Production 380 NTD x 330 Days = 125,400 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	-MCF	$.30 Avg.	$ -	$ -
	Coke, 3" Furnace Grade	2.08 NT	$24. Avg.	50.00	6,270.0
1.	Total Raw Materials	-	-	$ 50.00	$ 6,270.0
2.	Direct Labor	28,000 MH	$3.50	.78	98.0
3.	Supervision	-	$10,000	.08	10.0
4.	Maintenance	5% FC	-	8.17	1,025.0
5.	Plant Supplies	10%(L2+L3)	-	.08	10.8
	Utilities:				
6. a)	Electricity 26,100 KW	1,650KWH	$.01	16.50	2,070.0
b)	Cooling Water 1,160 gpm	4.4 MG	.05	.22	27.6
c)	Boiler Feed Water	MG	.25	-	-
d)	Fuel Gas	MCF	.30	-	-
7.	Total Utilities	-	-	16.72	2,097.6
8.	Other Direct (Catalyst)	10 NTY	$100.	.01	1.0
9.	TOTAL DIRECT	-	-	$ 76.85	$ 9,501.4
10.	Payroll Burden	30% L2	-	.24	29.4
11.	Plant Overhead	70%(L2+L3)	-	.60	75.6
12.	Pack & Ship	1000 MH	3.90	.06	7.8
13.	Waste Disposal	1000 MH	3.90	.06	7.8
14.	Other Indirect	-	-		
15.	TOTAL INDIRECT	-	-	$ .96	$ 120.6
16.	Depreciation	10% FC	-	16.30	2,050.0
17.	Taxes	2% FC	-	3.26	410.0
18.	Insurance	1% FC	-	1.63	205.0
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$ 21.19	$ 2,665.0
21.	TOTAL OPERATING	-	-	$ 97.99	$12,287.0
22.	Credits HP Steam	30.3 M Lb.	.50	15.15	1,900.0
23.	NET OPERATING	-	-	$ 82.84	$10,387.0

EXHIBIT 5-24b
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: Coke — CASE NO. 16 (Factored)
PROCESS: Lepsoe - Cominco — GAS NO. 2B

Fixed Capital, Battery Limits Basis $13,200,000
Sulfur Production 190 NTD x 330 Days = 62,700 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	-MCF	$.30 Avg.	$ -	$ -
	Coke, 3" Furnace Grade	2.08	$24. Avg.	50.00	3,135.0
1.	Total Raw Materials	-	-	$ 50.00	$3,135.0
2.	Direct Labor	28,000 MH	$3.50	1.56	98.0
3.	Supervision	1.	$10,000	.16	10.0
4.	Maintenance	5% FC	-	10.50	660.0
5.	Plant Supplies	10%(L2+L3)	-	.17	10.8
	Utilities:				
6. a)	Electricity 13,000 Kw	1,650 KWH	$.01	16.50	1,030.2
b)	Cooling Water 580 gpm	4.4 MG	.05	.22	13.8
c)	Boiler Feed Water	- MG	.25	-	-
d)	Fuel Gas	- MCF	.30	-	-
7.	Total Utilities	-	-	16.72	1,044.0
8.	Other Direct (Catalyst)	5 NTY	$100	.01	0.5
9.	TOTAL DIRECT	-	-	$ 79.12	$4,958.3
10.	Payroll Burden	30% L2	-	.46	29.4
11.	Plant Overhead	70%(L2+L3)	-	1.20	75.6
12.	Pack & Ship	1000 MH	3.90	.12	7.8
13.	Waste Disposal	1000 MH	3.90	.13	7.8
14.	Other Indirect	-	-	-	-
15.	TOTAL INDIRECT	-	-	$ 1.91	$ 120.6
16.	Depreciation	10% FC	-	21.05	1,320.0
17.	Taxes	2% FC	-	4.21	264.0
18.	Insurance	1% FC	-	2.10	132.0
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$ 27.36	$1,716.0
21.	TOTAL OPERATING	-	-	$108.39	$6,794.9
22.	Credits HP Steam	30.3 M Lb.	.50	15.15	950.0
23.	NET OPERATING	-	-	$ 93.24	$5,844.9

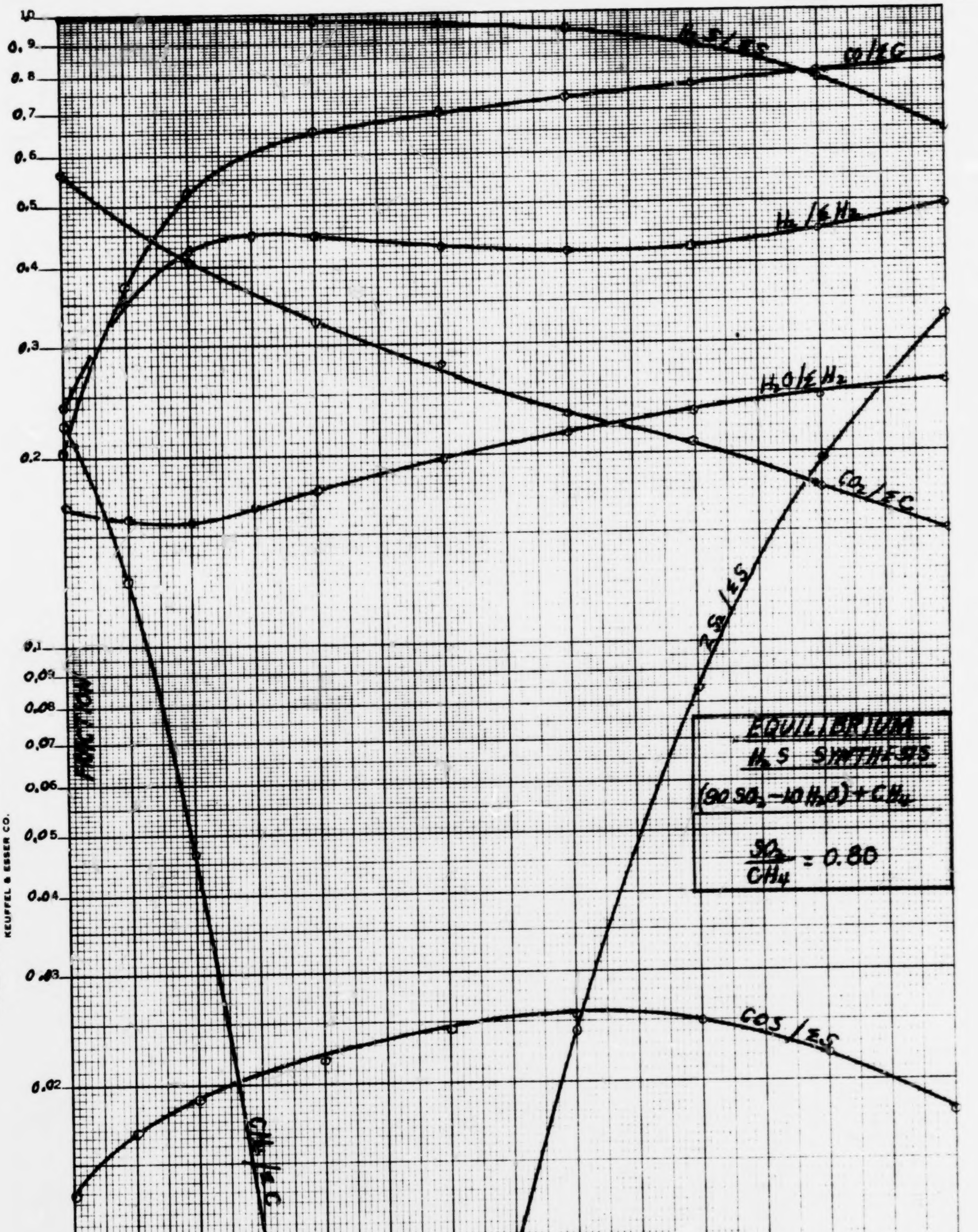
EXHIBIT 5-25

CASE 16

SUMMATION OF MAJOR EQUIPMENT

1.	Primary gas blower - multi-stage centrifugal blowers to handle 186,000 SCFM at an inlet temperature of 227°C with a Δp of 8 psi. Three units in parallel each equipped with an 8000 HP motor.	$950,000
2.	Reduction units - the size of these units has been held to 14 foot diameter to obtain satisfactory gas distribution and contact with coke. On this basis ten units will be required each consisting of a super-duty refractory lined vessel 14 foot I.D. with a coke depth of 12 feet, complete with coke feed mechanism rated at 4 1/2 tons per hour. Coke feed bin, double gate for pressure equalization and ash discharge system. Cost per unit - $200,000.	$2,000,000
3.	Combustion chamber - refractory lined carbon steel vessel with external insulation. Three units 12' diameter by 50 feet overall length with internal refractory baffles.	$450,000
4.	Waste heat boiler - a water tube type boiler will be installed to cool the process gas from 191,300 SCFM at 3,180°F to 890°F. This boiler will produce 630,000 lbs./hr. of 600 psig steam. The steam pressure is flexible and can be varied to suit end use conditions.	$1,100,000
5.	Hot gas precipitators - plate type precipitators of carbon steel construction with dust removal drag and screw conveyor system. Exact number of units to be installed in parallel has not been finally determined but total installation will handle 191,000 SCFM at 475°C.	$1,300,000
6.	Converter - eight units are required each to consist of a refractory lined steel vessel 30 feet in diameter packed to a depth of 6 feet with alumina type catalyst.	$680,000
7.	Sulfur condenser - eight units will be provided (one per converter) consisting of a horizontal waste heat boiler with integral steam drum operating at 50 psig inclined for sulfur drainage.	$640,000
	Total major purchased equipment	$7,620,000
	Total Fixed Capital	$22,860,000

EXHIBIT 6-1



K+E SEMI-LOGARITHMIC 46 5130
2 CYCLES X 140 DIVISIONS MADE IN U.S.A.
KEUFFEL & ESSER CO.

EXHIBIT 6-2

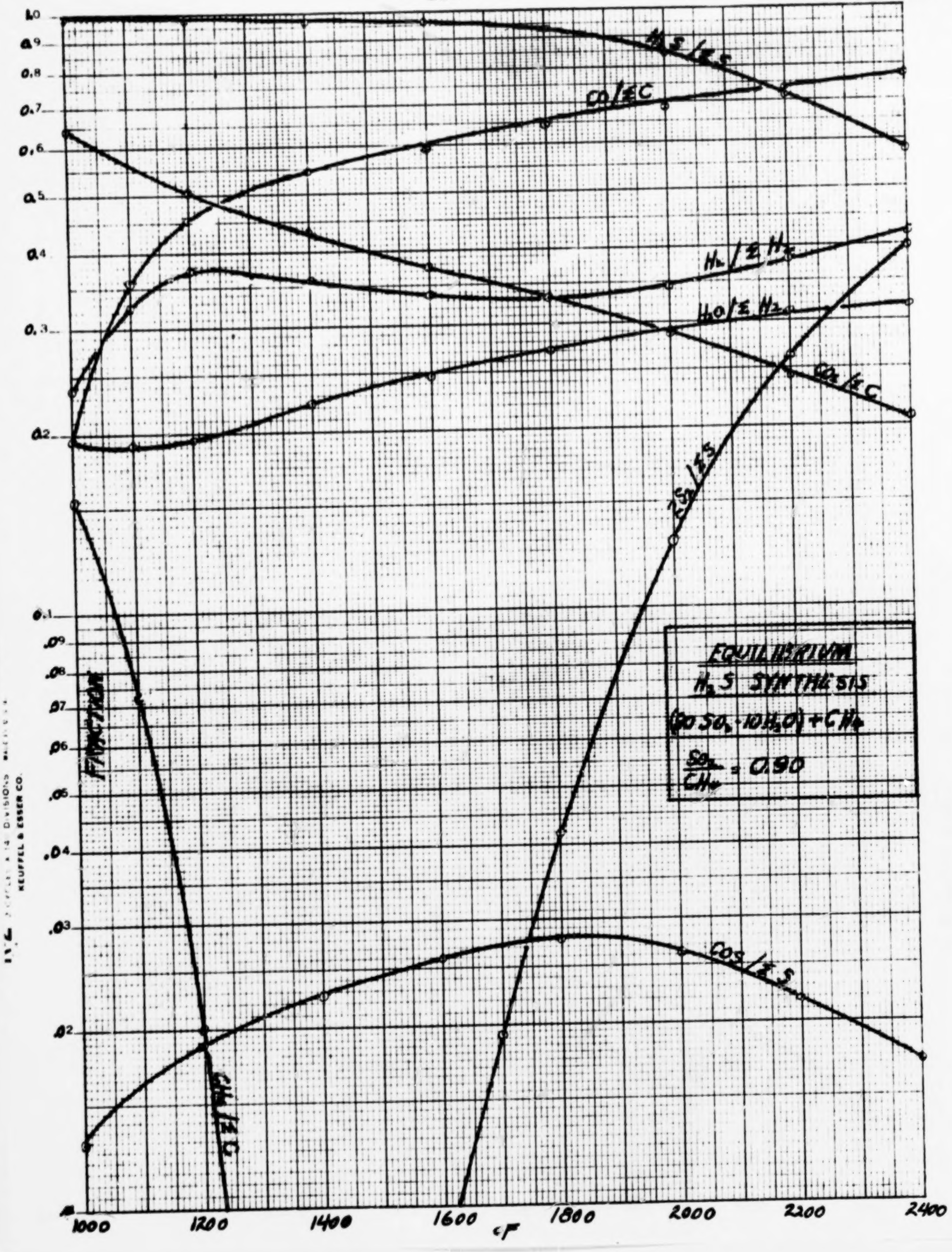

1.0
0.9
0.8
0.7
0.6
0.5
0.4
0.3
0.2
0.1
.09
.08
.07
.06
.05
.04
.03
.02
FRACTION
1000
1200
1400
1600
1800
2000
2200
2400
°F
H₂S/ΣS
CO/ΣC
H₂/ΣH₂
H₂O/ΣH₂
CO₂/ΣC
ΣS₂/ΣS
COS/ΣS
CH₄/ΣC
EQUILIBRIUM
H₂S SYNTHESIS
(90 SO₂-10 H₂O) + CH₄
SO₂/CH₄ = 0.90
KEUFFEL & ESSER CO.

EXHIBIT 6-3

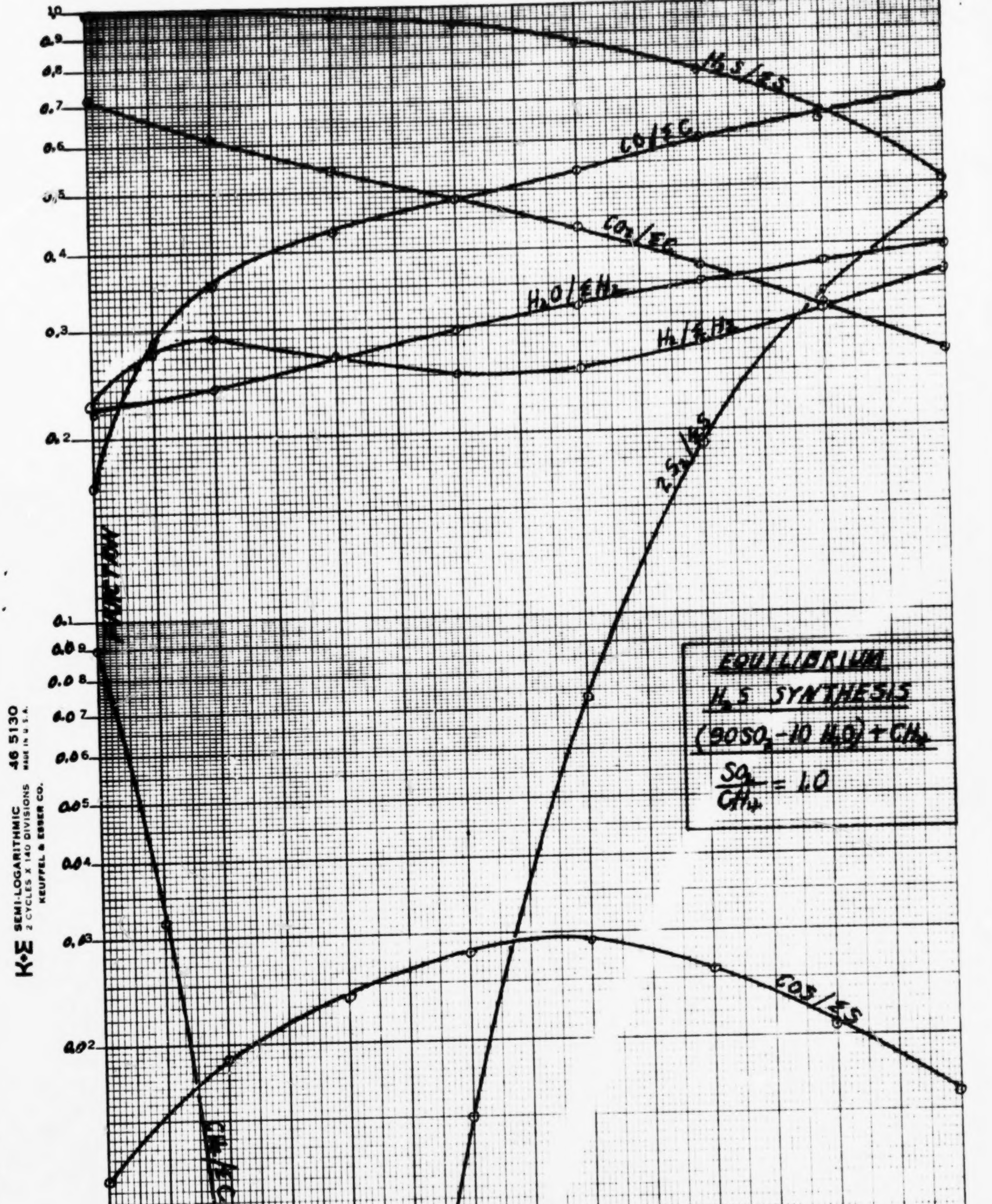

1.0
0.9
0.8
0.7
0.6
0.5
0.4
0.3
0.2
0.1
0.09
0.08
0.07
0.06
0.05
0.04
0.03
0.02
0.01
1000
1200
1400
1600
1800
2000
2200
2400
°F
FRACTION
$H_2S/\Sigma S$
$CO/\Sigma C$
$CO_2/\Sigma C$
$H_2O/\Sigma H_2$
$H_2/\Sigma H_2$
$2S_2/\Sigma S$
$COS/\Sigma S$
$CH_4/\Sigma C$
EQUILIBRIUM
H_2S SYNTHESIS
$(90SO_2-10\ H_2O) + CH_4$
$\frac{SO_2}{CH_4} = 1.0$
K+E SEMI-LOGARITHMIC 2 CYCLES X 140 DIVISIONS KEUFFEL & ESSER CO.
46 5130 MADE IN U.S.A.

EXHIBIT 6-4

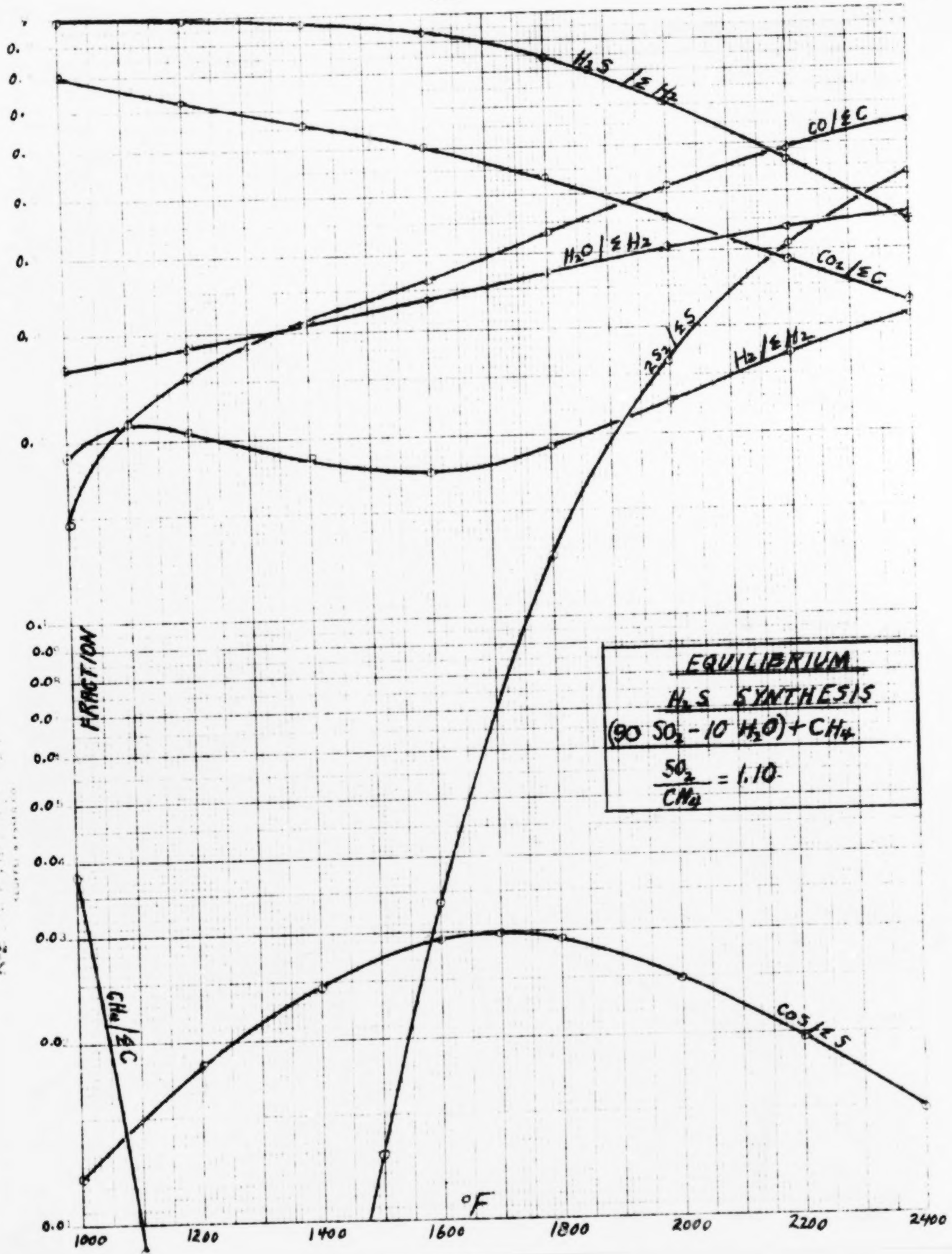
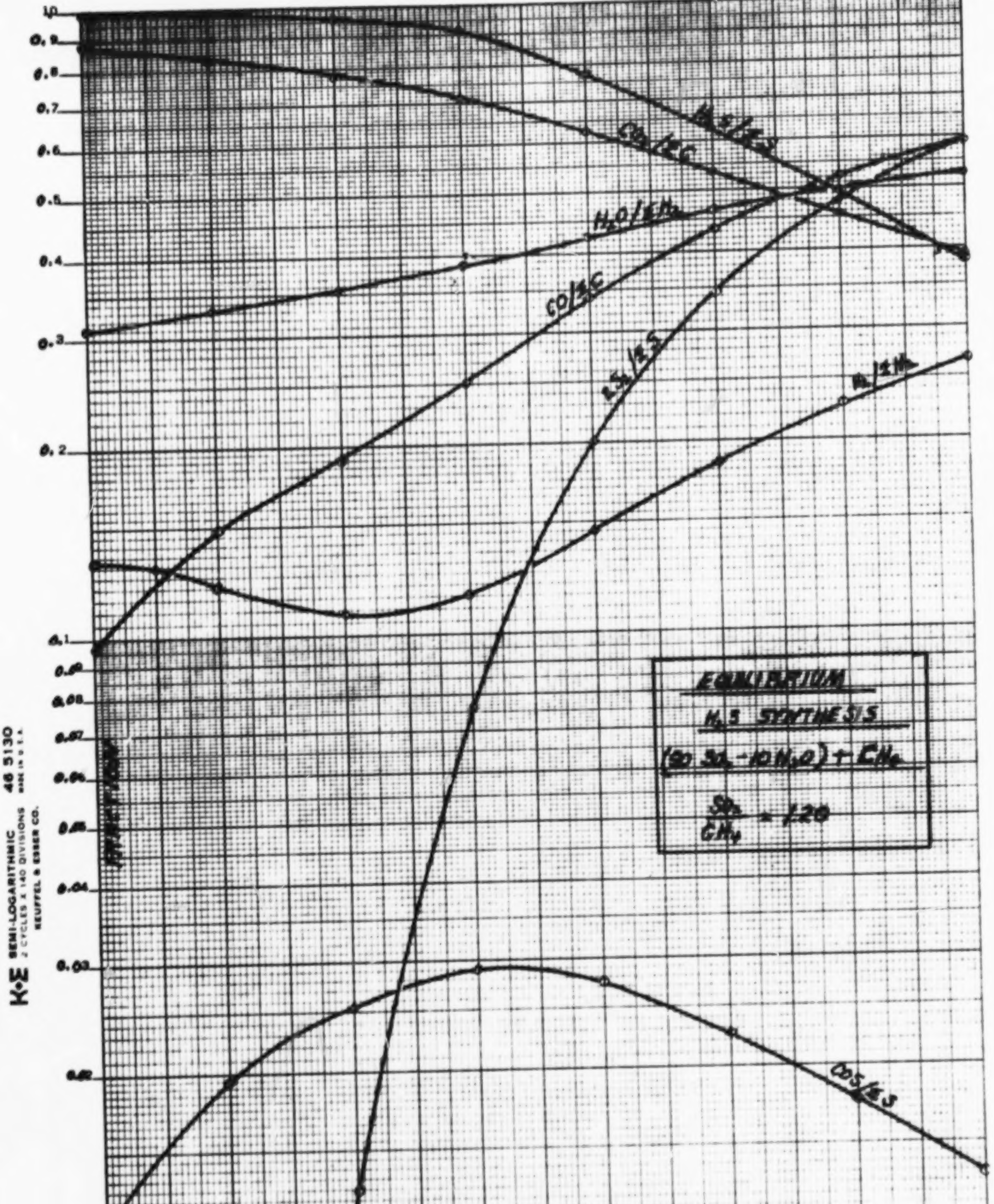
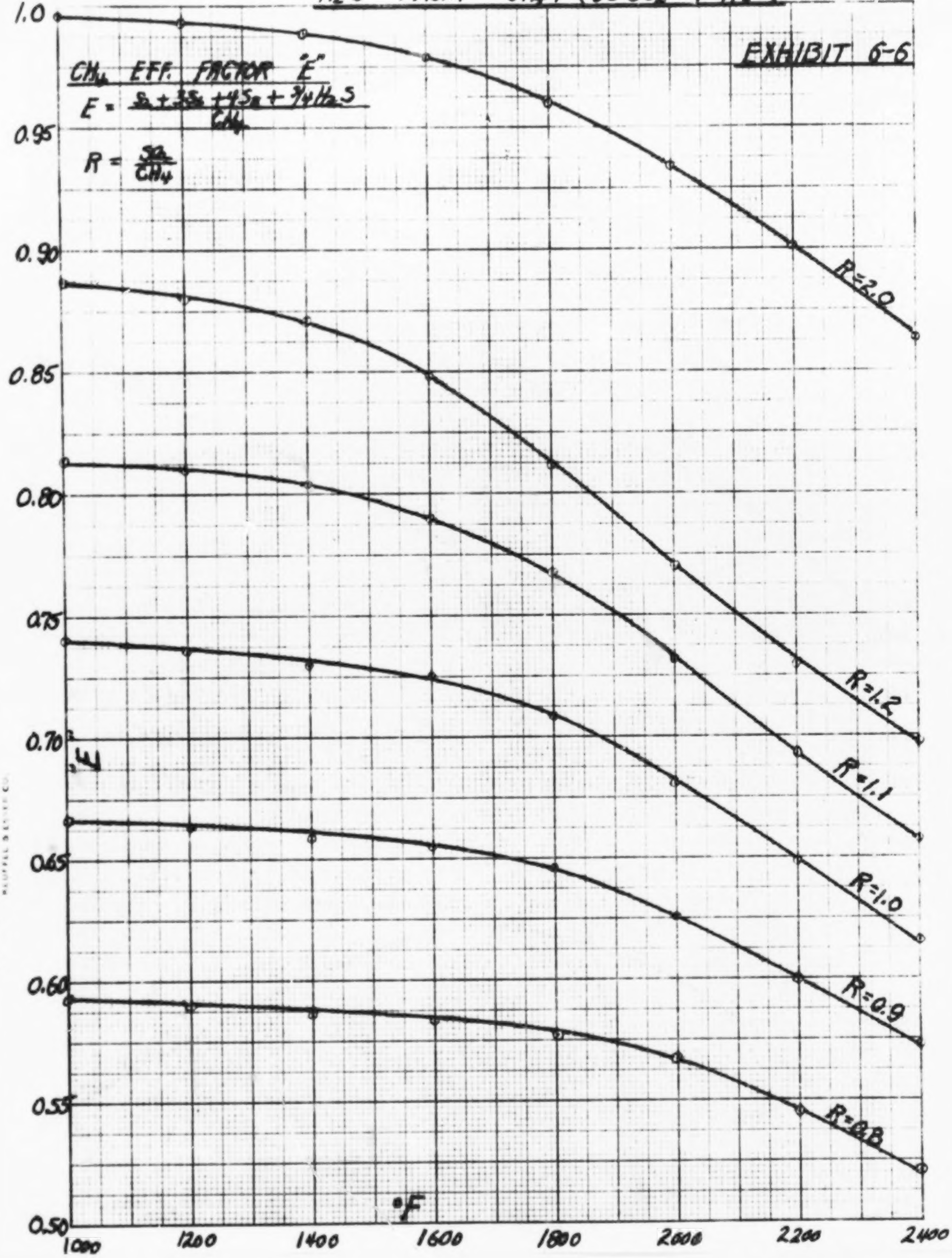

H2S / ΣH2
CO / ΣC
CO2 / ΣC
H2O / ΣH2
2S2 / ΣS
H2 / ΣH2
COS / ΣS
CH4 / ΣC
FRACTION
°F
1000
1200
1400
1600
1800
2000
2200
2400
0.03
0.0
0.01
EQUILIBRIUM
H2S SYNTHESIS
(90 SO2 - 10 H2O) + CH4
SO2 / CH4 = 1.10

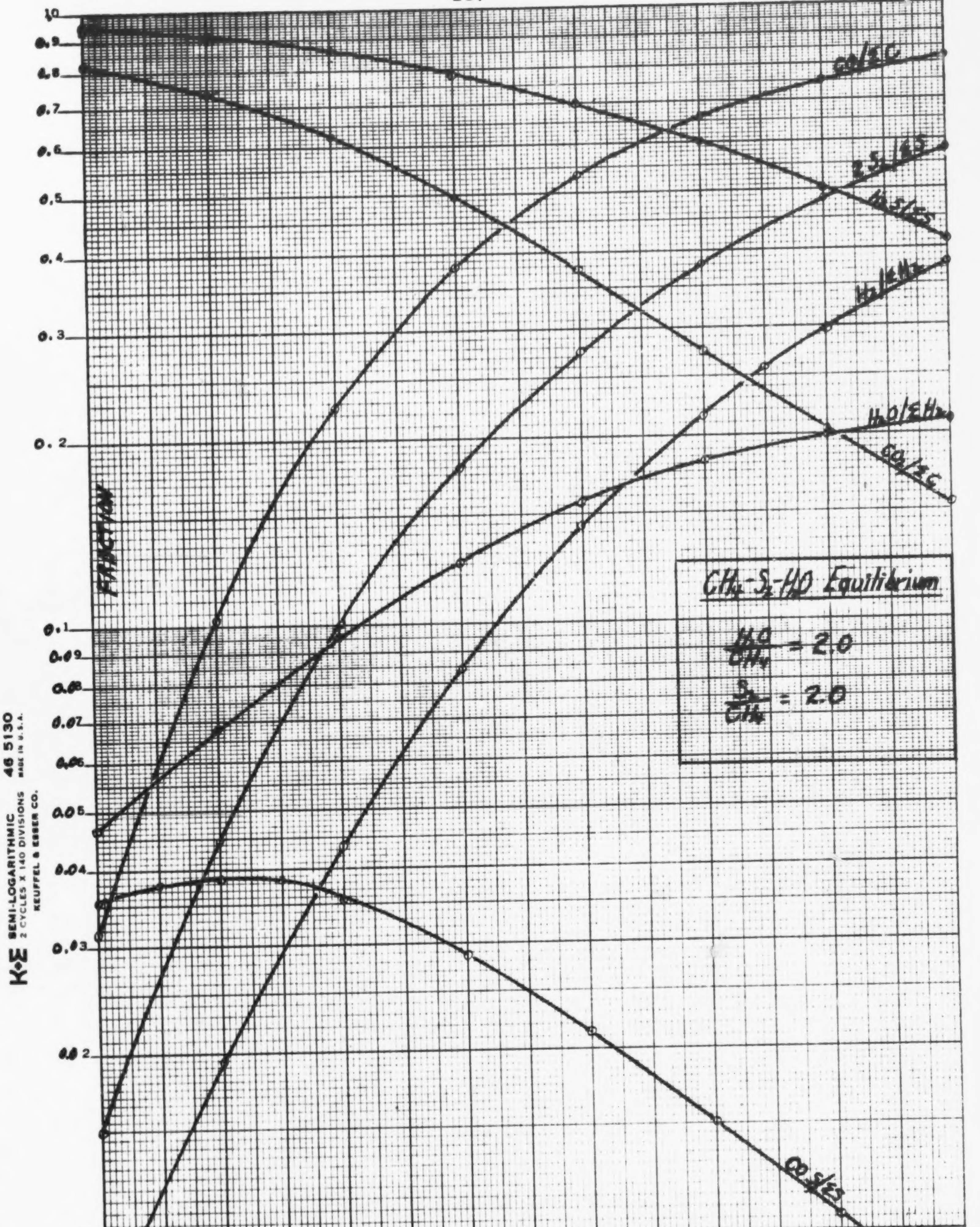
EXHIBIT 6-5


1.0
0.9
0.8
0.7
0.6
0.5
0.4
0.3
0.2
0.1
0.09
0.08
0.07
0.06
0.05
0.04
0.03
0.02
0.01
1000
1200
1400
1600
1800
2000
2200
2400
°F
$H_2S/\Sigma S$
$CO_2/\Sigma C$
$H_2O/\Sigma H_2$
$CO/\Sigma C$
$S_2/\Sigma S$
$H_2/\Sigma H_2$
$COS/\Sigma S$
EQUILIBRIUM
H_2S SYNTHESIS
$(90\ SO_2 - 10\ H_2O) + CH_4$
$\frac{SO_2}{CH_4} = 1.20$
K+E SEMI-LOGARITHMIC 2 CYCLES X 140 DIVISIONS KEUFFEL & ESSER CO. 46 5130 MADE IN U.S.A.

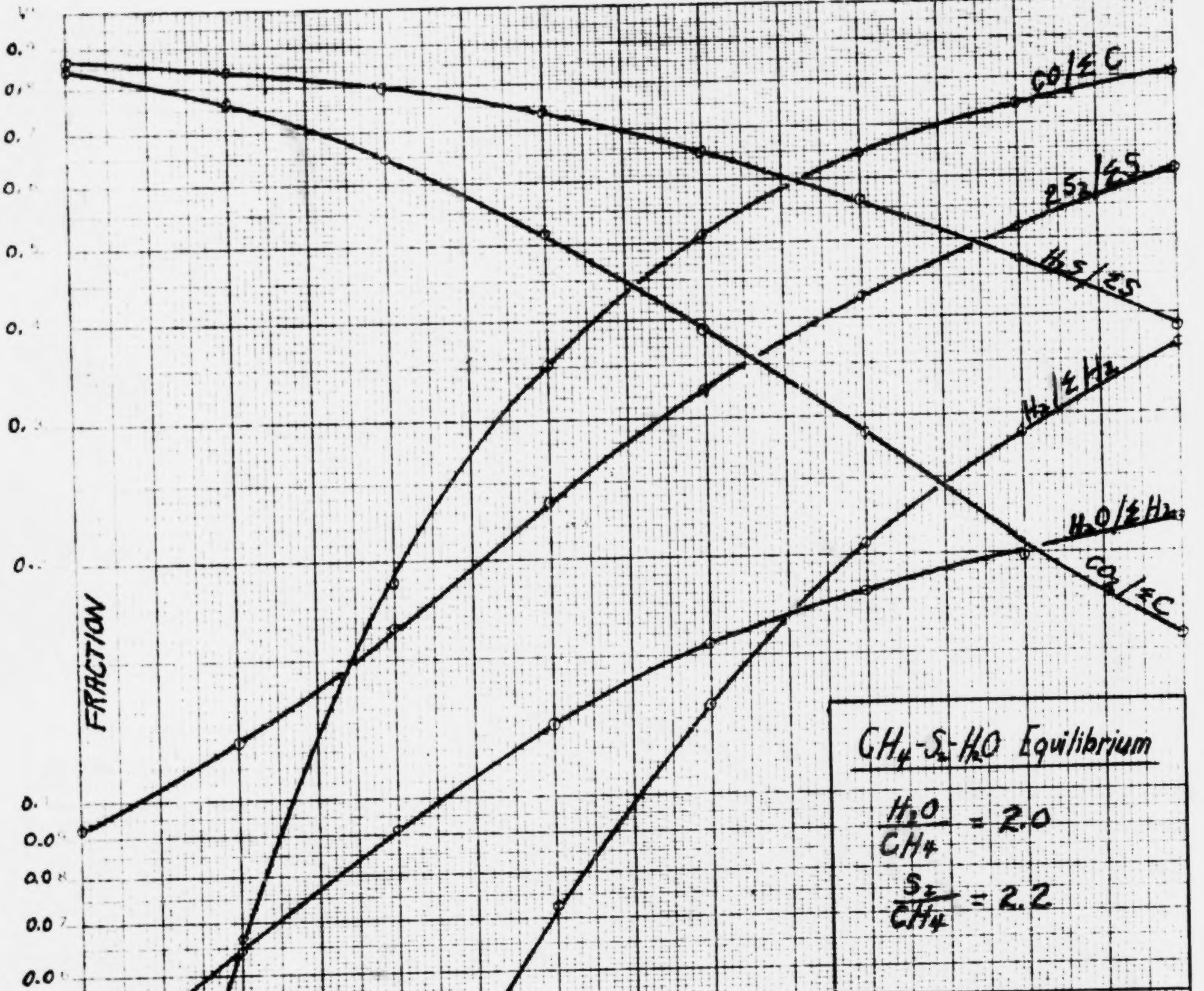
H_2S FROM $CH_4 + (90 SO_2 - 10 H_2O)$



EXH 6-7


CH4-S2-H2O Equilibrium
H2O/CH4 = 2.0
S2/CH4 = 2.0
FRACTION
°F
CO/ΣC
2S2/ΣS
H2S/ΣS
H2/ΣH2
H2O/ΣH2
CO2/ΣC
COS/ΣS
1.0
0.9
0.8
0.7
0.6
0.5
0.4
0.3
0.2
0.1
0.09
0.08
0.07
0.06
0.05
0.04
0.03
0.02
0.01
1000
1200
1400
1600
1800
2000
2200
2400
K+E SEMI-LOGARITHMIC 46 5130
2 CYCLES X 140 DIVISIONS MADE IN U.S.A.
KEUFFEL & ESSER CO.

EXH. 6-8

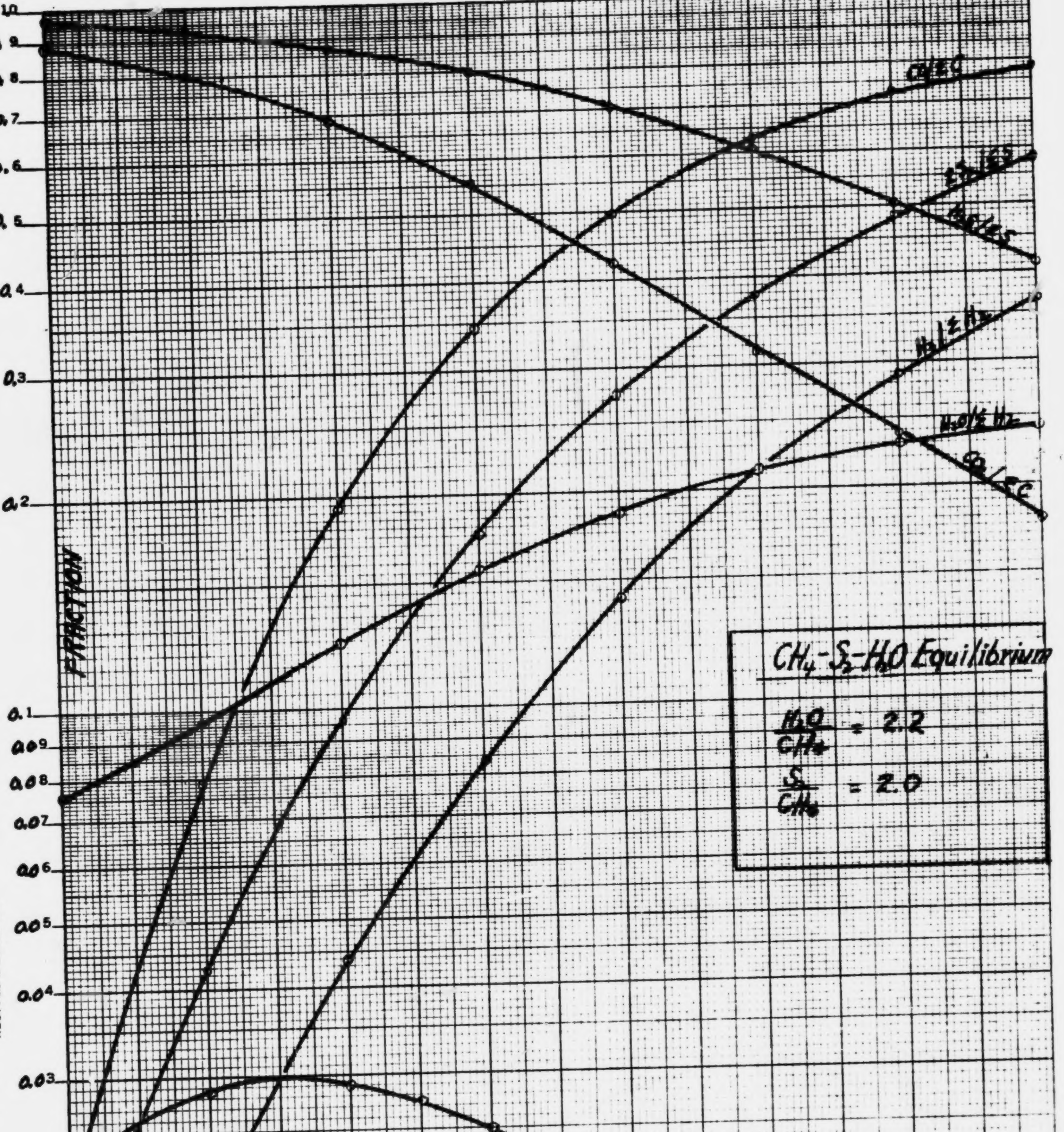


CH_4-S_2-H_2O Equilibrium

$\frac{H_2O}{CH_4} = 2.0$

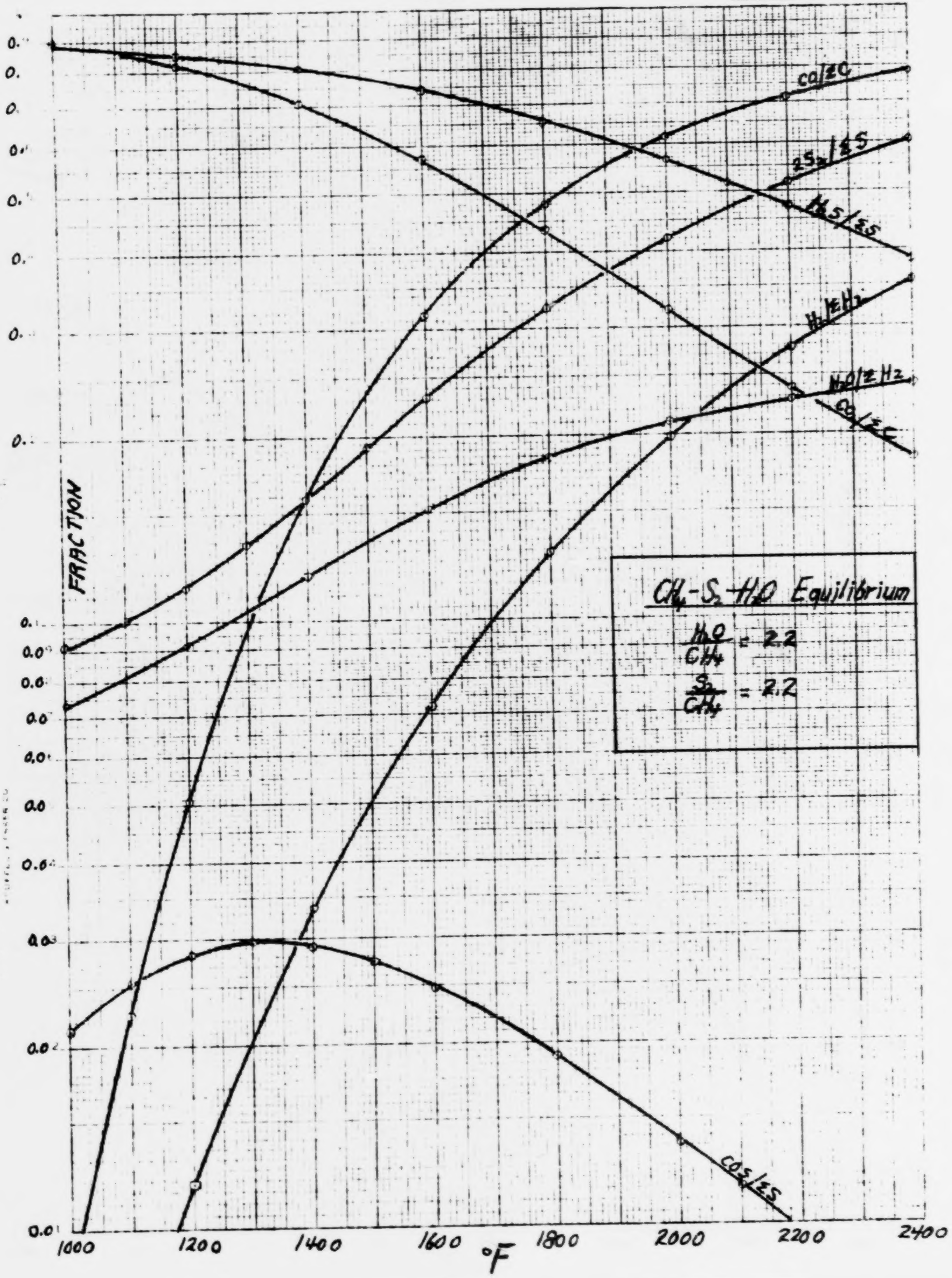
$\frac{S_2}{CH_4} = 2.2$

EXH. 6-9

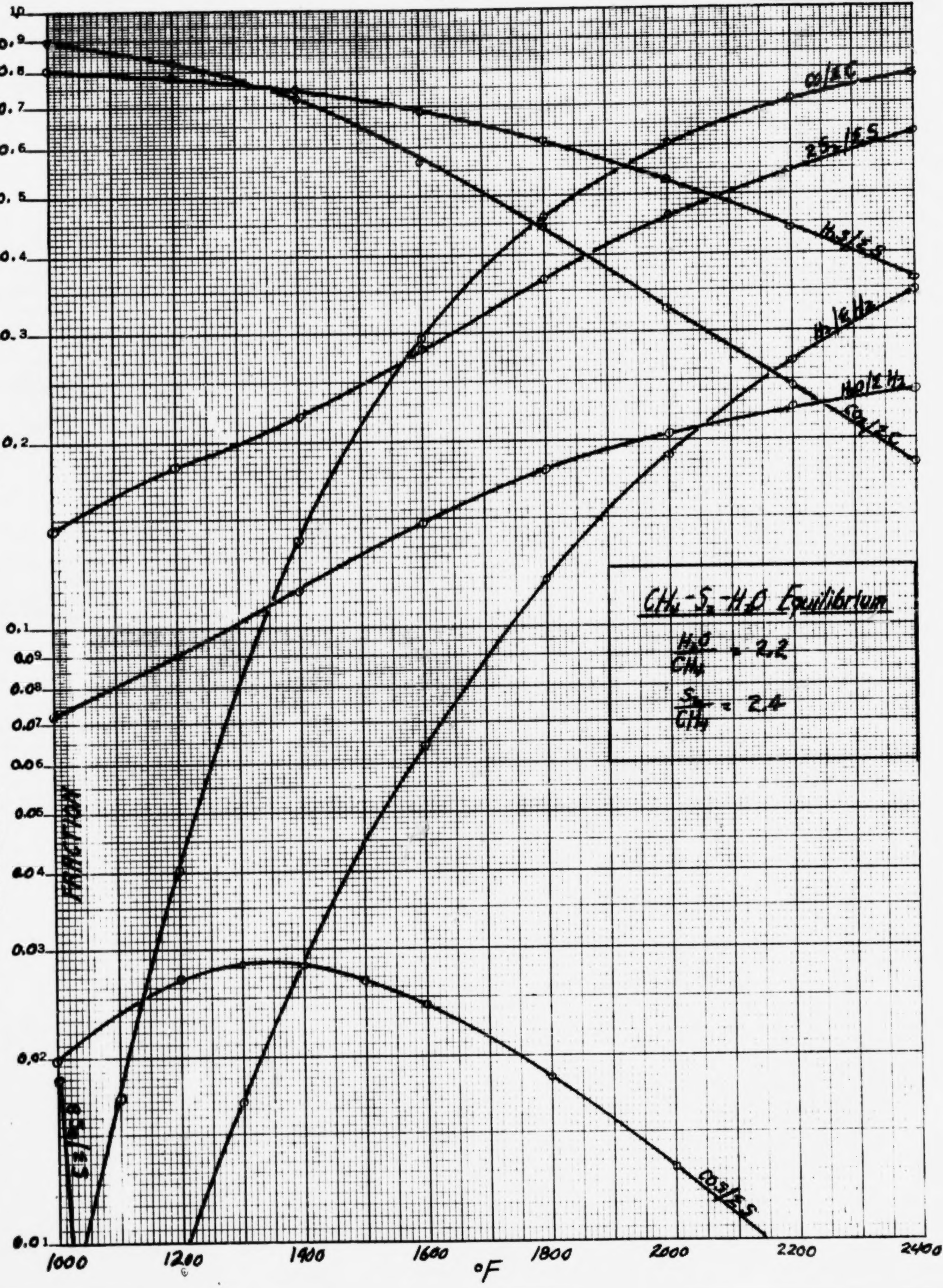


K+E SEMI-LOGARITHMIC 46 5130
2 CYCLES X 140 DIVISIONS MADE IN U.S.A.
KEUFFEL & ESSER CO.

EXH 6-10


CH_4-S_2-H_2O Equilibrium
$\frac{H_2O}{CH_4}$ = 2.2
$\frac{S_2}{CH_4}$ = 2.2
CO/ΣC
$2S_2$/ΣS
H_2S/ΣS
H_2/ΣH_2
H_2O/ΣH_2
CO_2/ΣC
COS/ΣS
FRACTION
0.03
0.02
0.01
1000
1200
1400
1600
1800
2000
2200
2400
°F

EXH 6-11



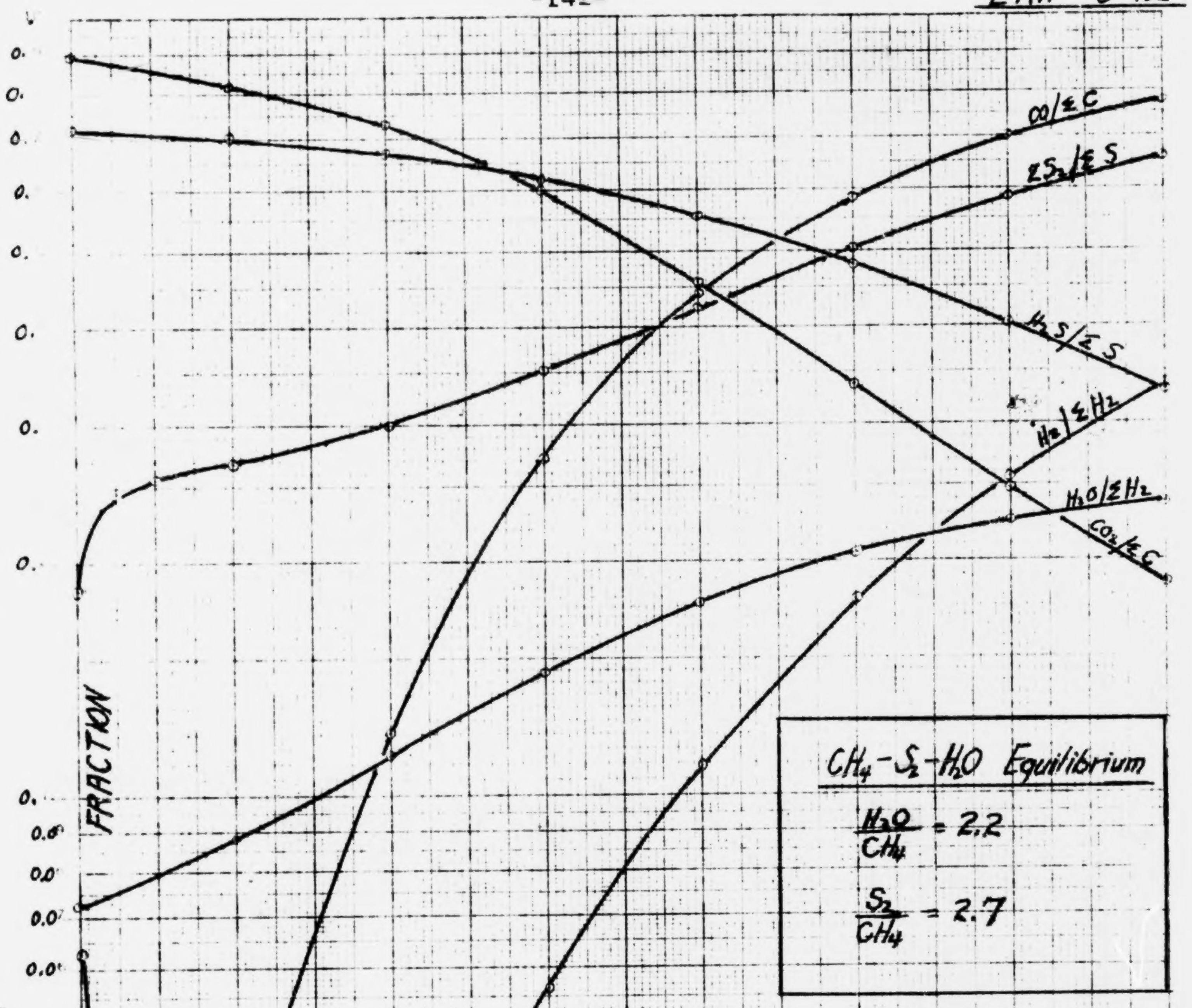
CH_4-S_2-H_2O Equilibrium

$\frac{H_2O}{CH_4} = 2.2$

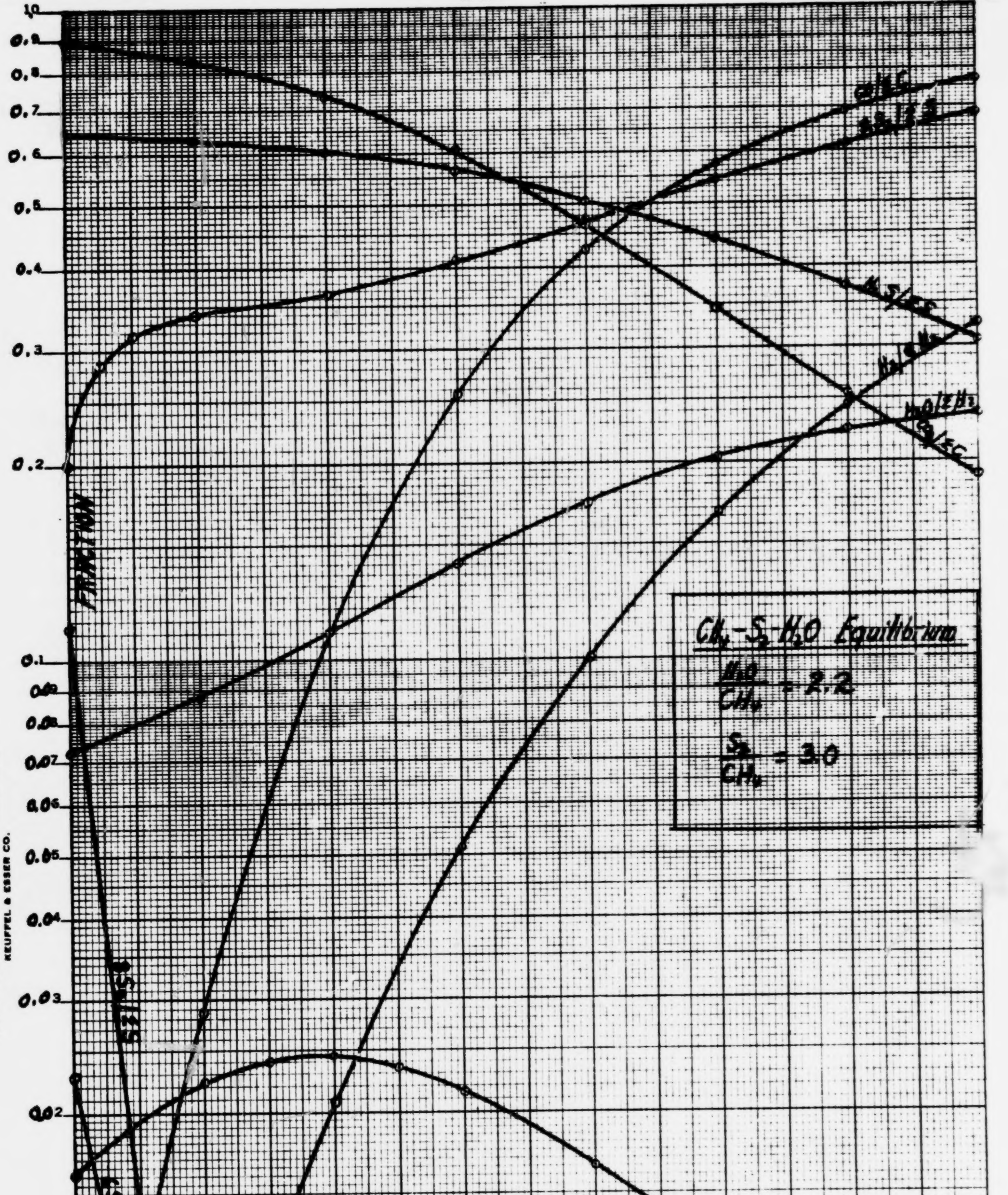
$\frac{S_2}{CH_4} = 2.4$

K+E SEMI-LOGARITHMIC 46 5130
2 CYCLES X 140 DIVISIONS MADE IN U.S.A.
KEUFFEL & ESSER CO.

EXH 6-12

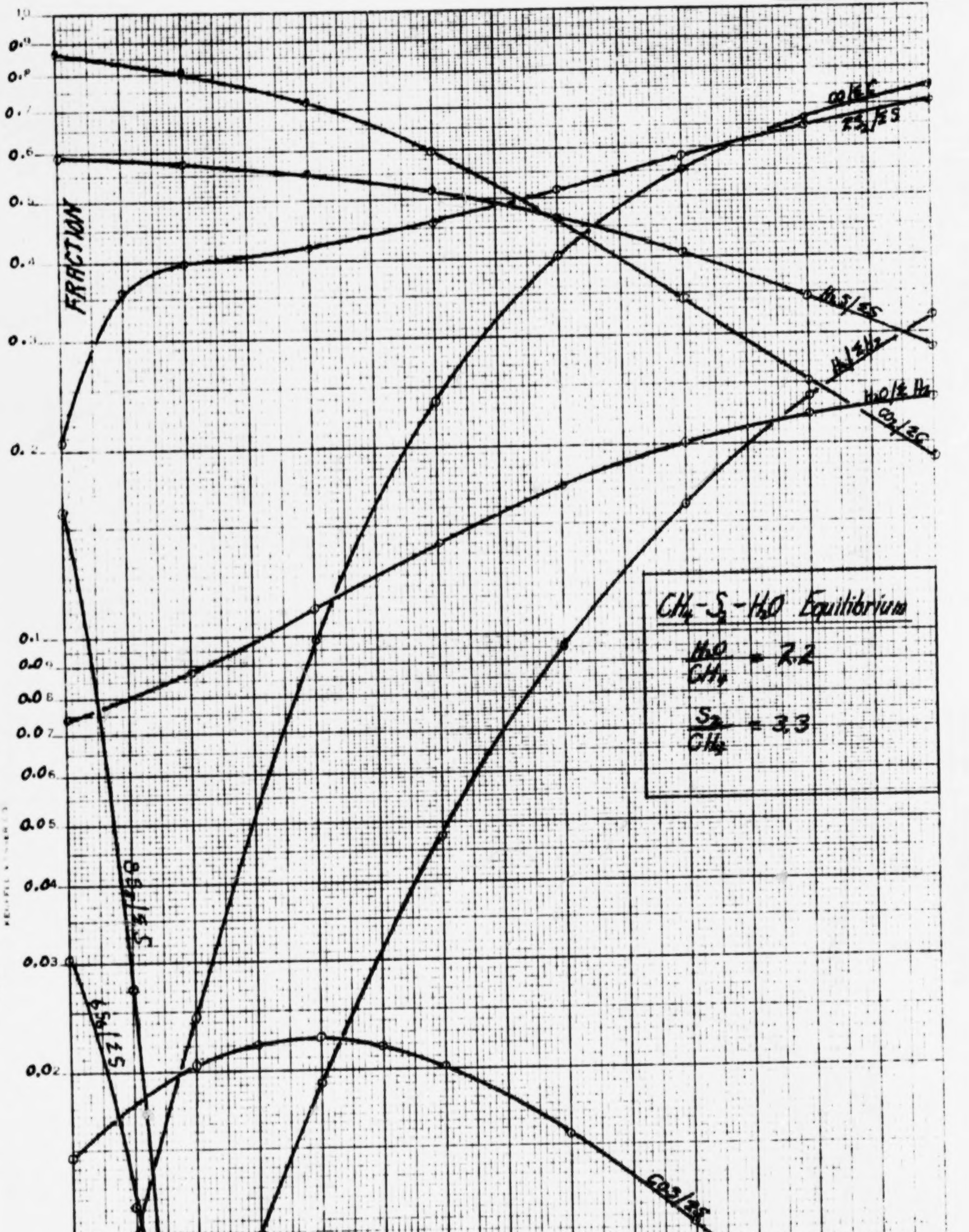

CO/ΣC
ΣS₂/ΣS
H₂S/ΣS
H₂/ΣH₂
H₂O/ΣH₂
CO₂/ΣC
COS/ΣS
FRACTION
CH₄-S₂-H₂O Equilibrium
H₂O/CH₄ = 2.2
S₂/CH₄ = 2.7
0.02
0.01
1000
1200
1400
1600
1800
2000
2200
2400
°F

EXH 6-13

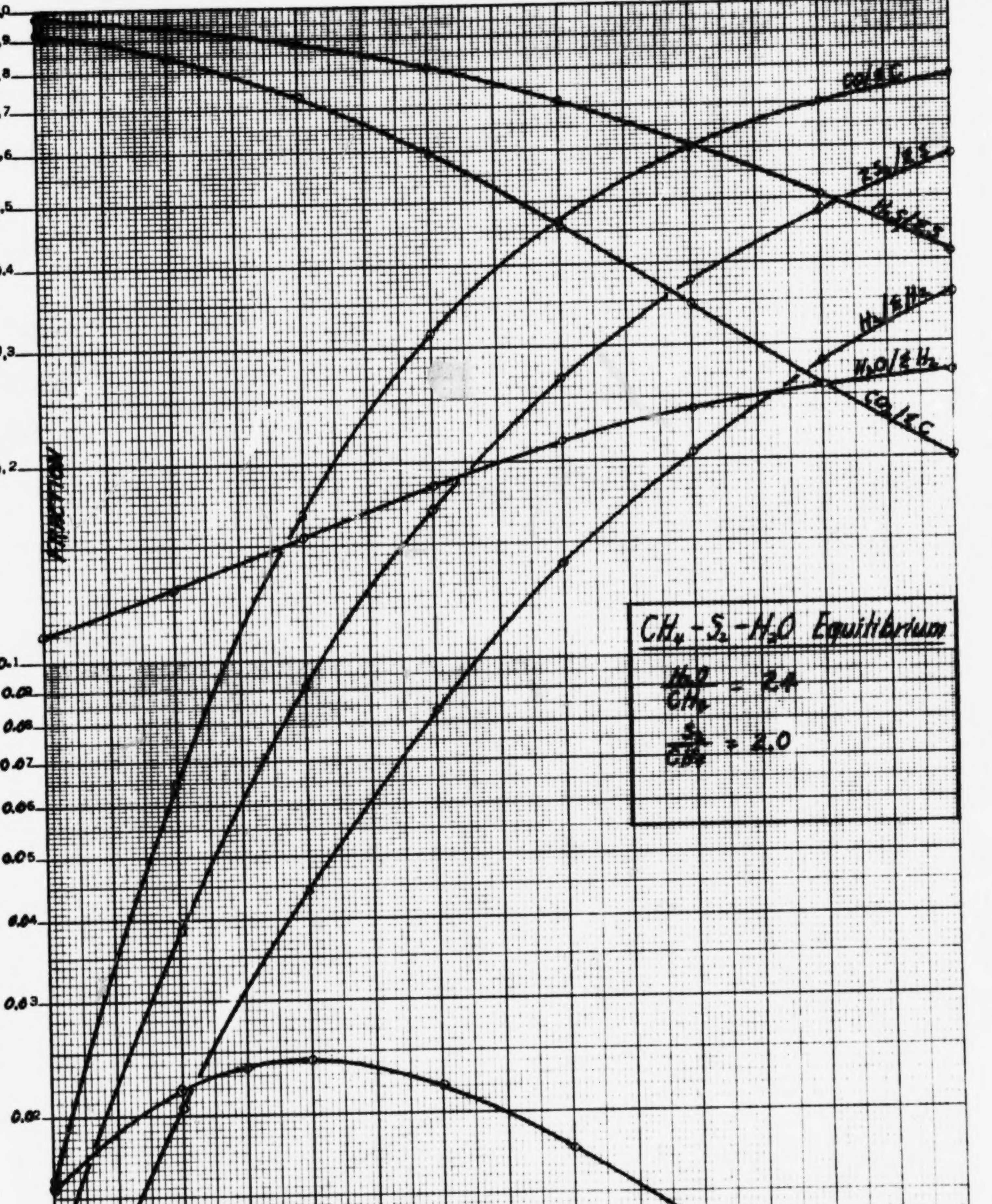


K+E SEMI-LOGARITHMIC 46 5130
2 CYCLES X 140 DIVISIONS MADE IN U.S.A.
KEUFFEL & ESSER CO.

EXH 6-14

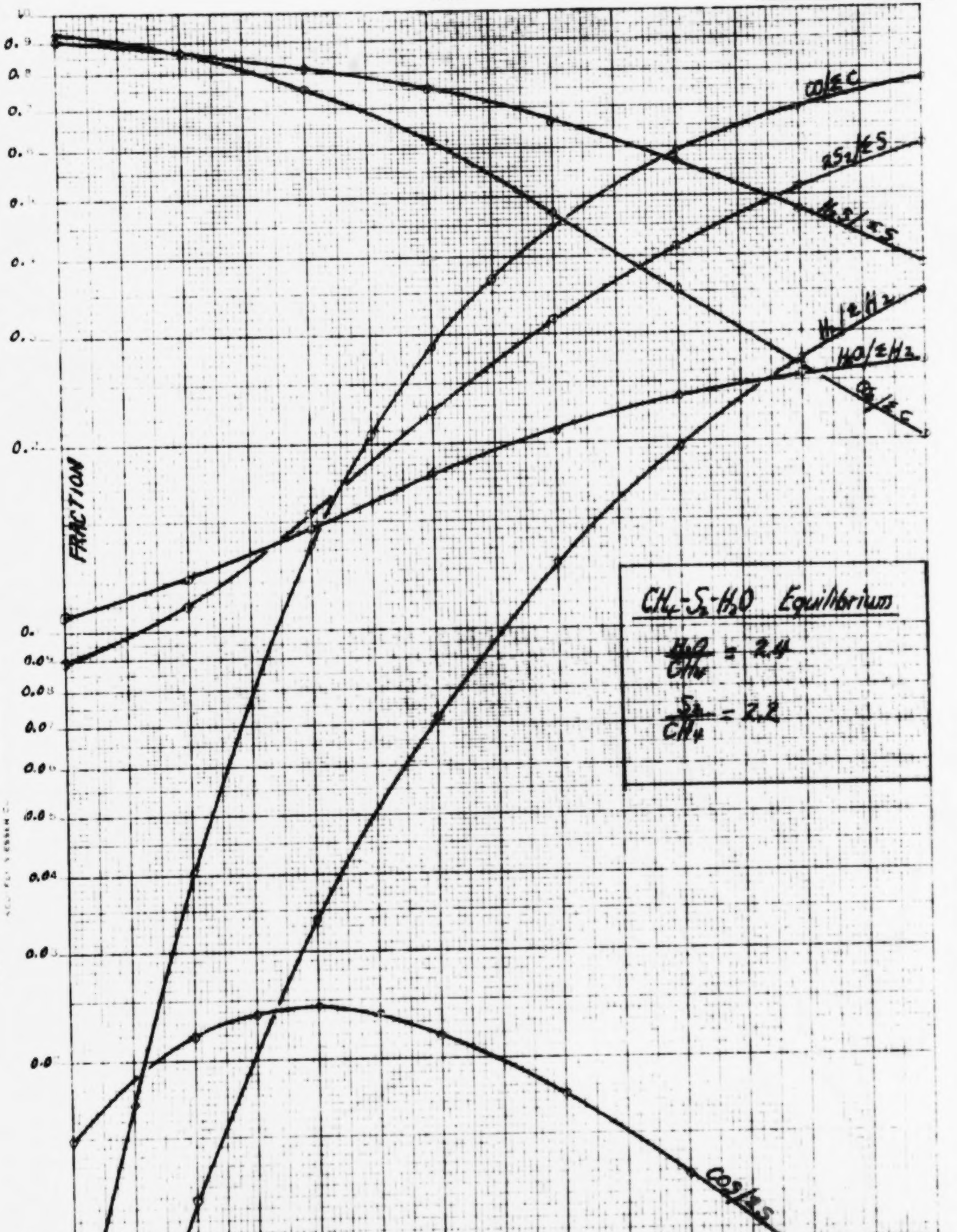

FRACTION
1.0
0.9
0.8
0.7
0.6
0.5
0.4
0.3
0.2
0.1
0.09
0.08
0.07
0.06
0.05
0.04
0.03
0.02
0.01
1000
1200
1400
1600
1800
2000
2200
2400
°F
CO/ΣC
ΣS₂/ΣS
H₂S/ΣS
H₂/ΣH₂
H₂O/ΣH₂
CO₂/ΣC
COS/ΣS
CS₂/ΣC
CS₂/ΣS
CH₄-S₂-H₂O Equilibrium
H₂O/CH₄ = 7.2
S₂/CH₄ = 3.3
K&E SEMI-LOGARITHMIC 46 5130

EXH 6-15

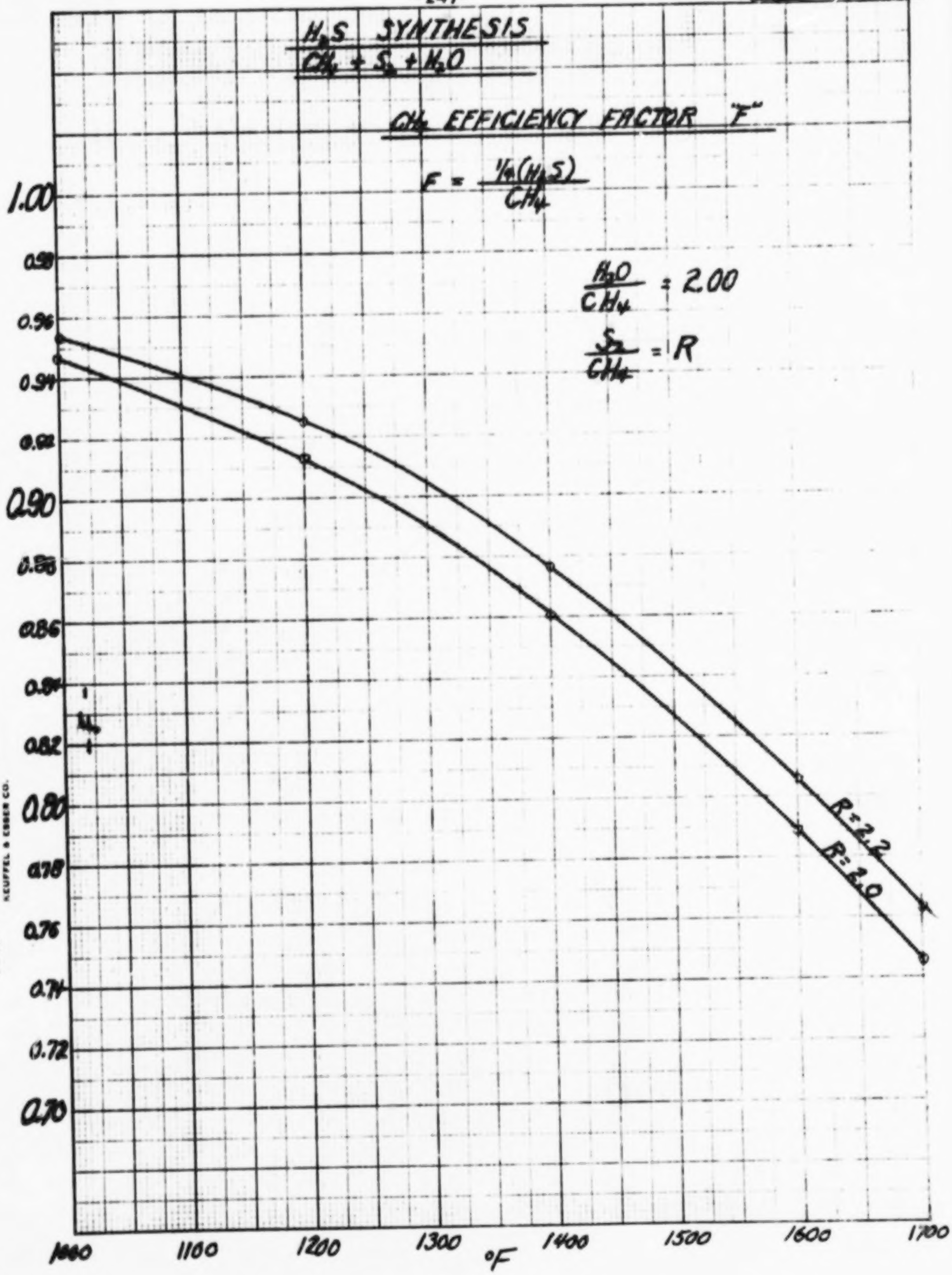


K&E SEMI-LOGARITHMIC 46 5130
2 CYCLES X 140 DIVISIONS MADE IN U.S.A.
KEUFFEL & ESSER CO.

EXH 6-16


FRACTION
1.0
0.9
0.8
0.7
0.6
0.5
0.4
0.3
0.2
0.1
0.09
0.08
0.07
0.06
0.05
0.04
0.03
0.02
0.01
1000
1200
1400
1600
1800
2000
2200
2400
°F
CO/ΣC
2S₂/ΣS
H₂S/ΣS
H₂/ΣH₂
H₂O/ΣH₂
CO₂/ΣC
COS/ΣS
CH₄-S₂-H₂O Equilibrium
H₂O/CH₄ = 2.4
S₂/CH₄ = 2.2

EXH 6-17



H_2S SYNTHESIS

$CH_4 + S_2 + H_2O$

CH_4 EFFICIENCY FACTOR "F"

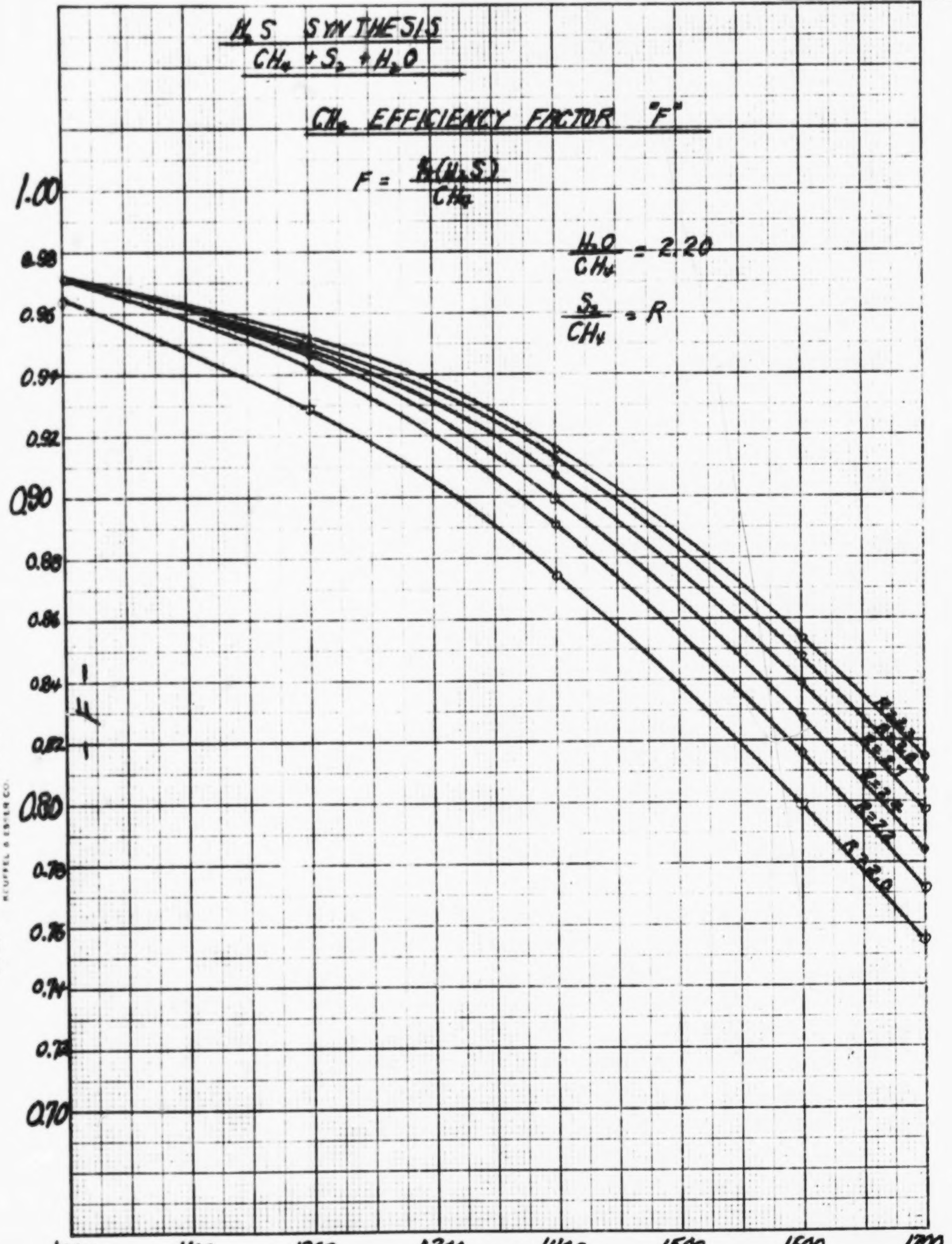
$$F = \frac{1/4(H_2S)}{CH_4}$$

$$\frac{H_2O}{CH_4} = 2.00$$

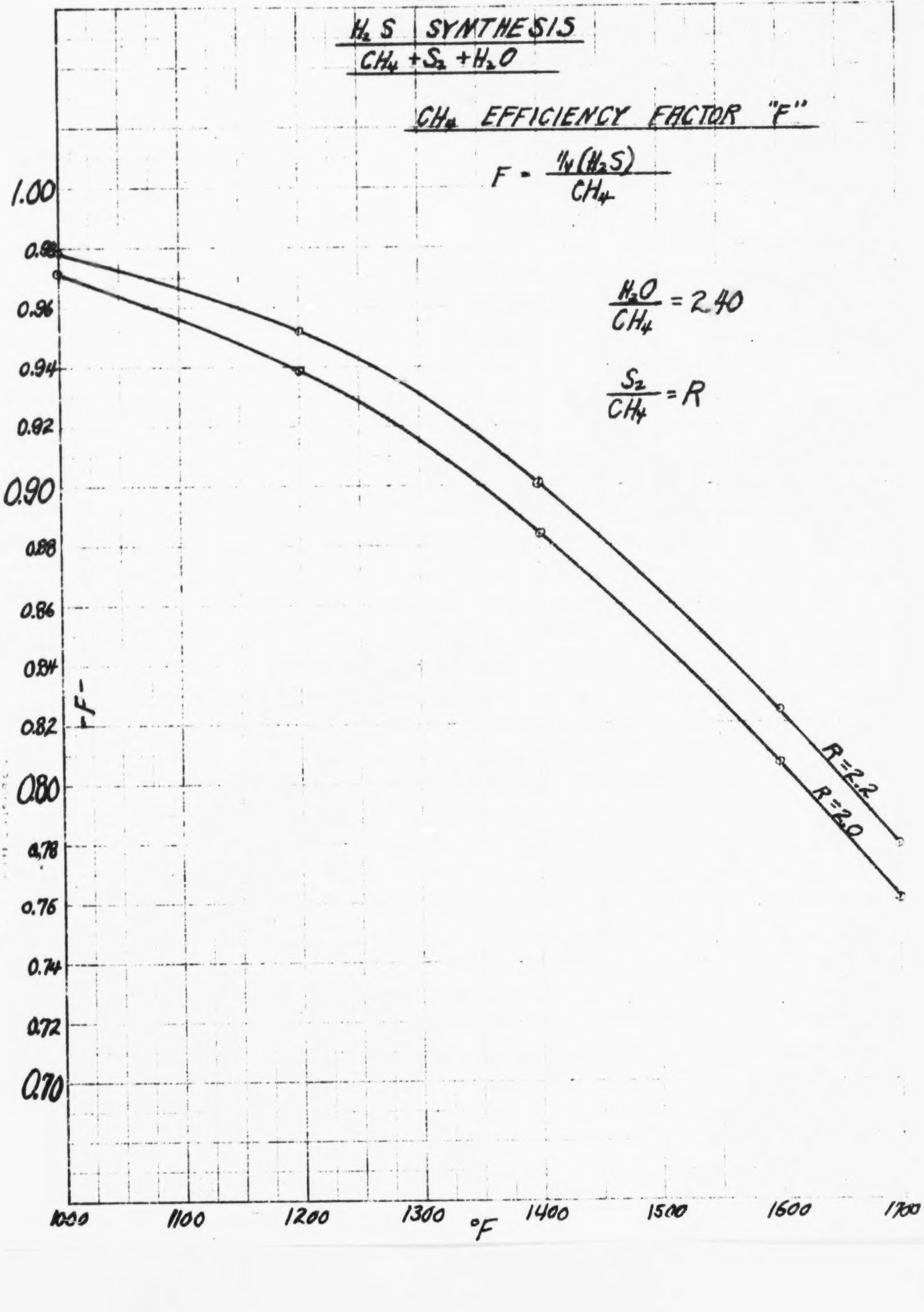
$$\frac{S_2}{CH_4} = R$$

K&E 20 X 20 TO THE INCH 46 1240
KEUFFEL & ESSER CO.

EXH 6-18


H_2S SYNTHESIS
$CH_4 + S_2 + H_2O$
CH_4 EFFICIENCY FACTOR "F"
$F = \frac{\frac{1}{2}(H_2S)}{CH_4}$
$\frac{H_2O}{CH_4} = 2.20$
$\frac{S_2}{CH_4} = R$
1.00
0.98
0.96
0.94
0.92
0.90
0.88
0.86
0.84
0.82
0.80
0.78
0.76
0.74
0.72
0.70
R=2.4
R=2.2
R=2.0
1000
1100
1200
1300
1400
1500
1600
1700
°F

EXH G-19



H$_2$S SYMTHESIS

$CH_4 + S_2 + H_2O$

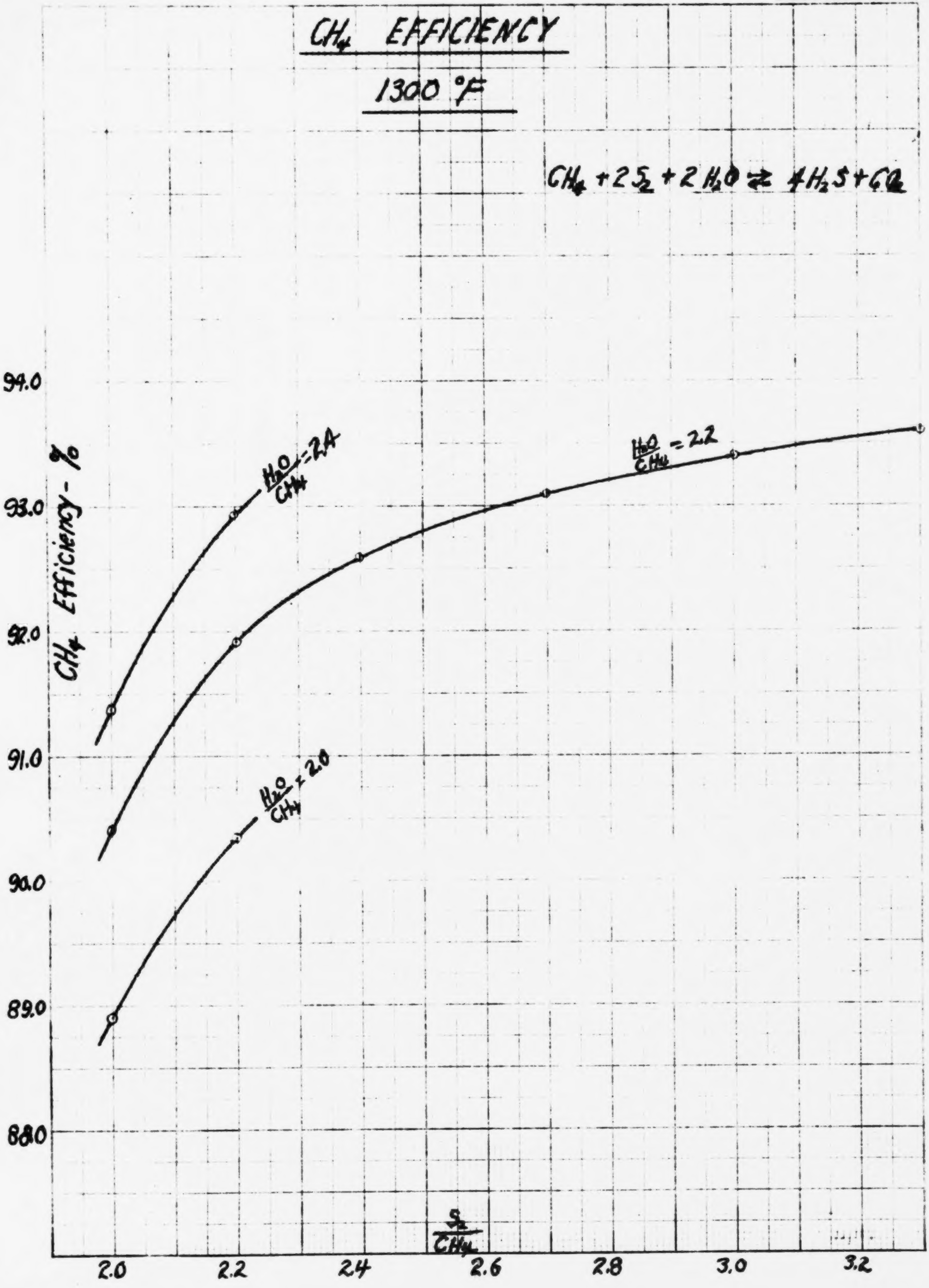
CH_4 EFFICIENCY FACTOR "F"

$$F = \frac{\frac{1}{4}(H_2S)}{CH_4}$$

$$\frac{H_2O}{CH_4} = 2.40$$

$$\frac{S_2}{CH_4} = R$$

EXH 6-20


CH4 EFFICIENCY
1300 °F
CH4 + 2 S2 + 2 H2O ⇄ 4 H2S + CO2
CH4 Efficiency - %
94.0
93.0
92.0
91.0
90.0
89.0
88.0
H2O/CH4 = 2.4
H2O/CH4 = 2.2
H2O/CH4 = 2.0
S2/CH4
2.0
2.2
2.4
2.6
2.8
3.0
3.2

H_2S REDUCTION OF SO_2

EXH. 6-21A

CONVERTER GAS 2A 20000 SCFM

CASE 6-1

PROCESS POINTS	A	B	C	D	E	F	H	J	K	L	V	W
TEMP °C	25	25	625	25	25	625	625	625	720	373	150	
VOL. SCFM	2760	880	1870	8610	20000	2090	12960	16070	17210	37220	33680	38900
VOL CFM @T°	3010	960	6150	9400	21830	6870	42630	52850	62600	88070	52183	
% CH_4	100.0	100.0	100.0					11.10				
% H_2O						100.00	13.14	22.20	4.43	2.07	20.72	18.39
% H_2S									37.58	17.38	0.77	
% CO_2							6.58	5.06	11.44	5.29	6.79	6.99
% COS									1.94	0.90	0.04	
% CO									2.39	1.11	1.22	
% S_2							28.90	22.20	1.94	0.90	0.19	
% H_2									1.49	0.69	0.76	
% CS_2									0.08	0.04	0.04	
% SO_2					15.80					8.49	0.41	1.69
% O_2				21.0	1.20					0.64		0.95
% N_2				79.0	83.00		51.38	39.44	38.68	62.51	69.06	71.98

SULFUR YIELD - 82.16 %

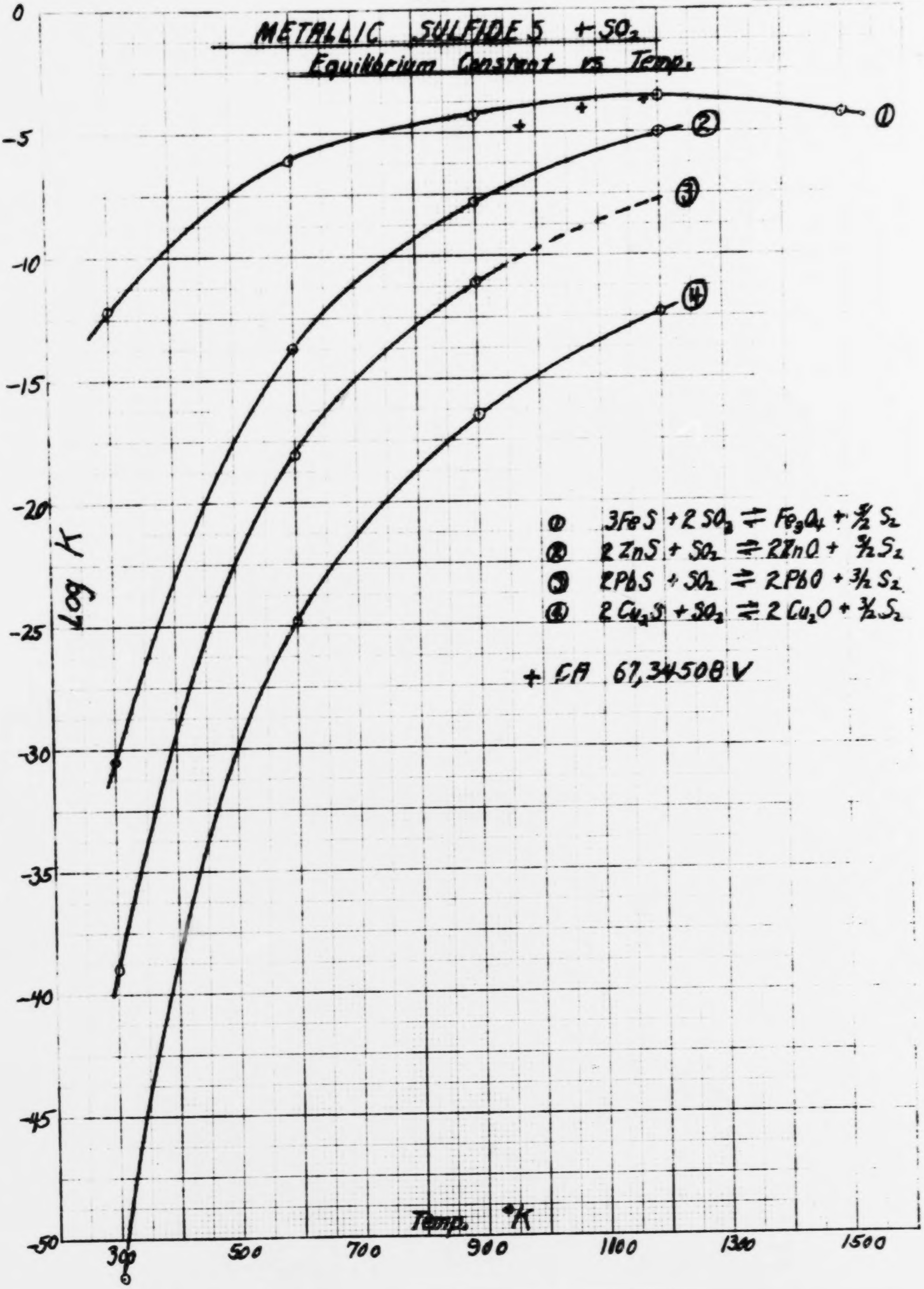
METHANE FACTOR - 24060 SCFM/TON S

SULFUR DISTRIBUTION

		#/min	NT/Day	Poin
GROSS	-	898.1	646.6	Ⓜ
RECYCLE	-	667.0	480.2	Ⓖ
NET PRODUCT	-	231.1	166.4	Ⓝ

EXHIBIT 6-21A

EXH 6-22



METALLIC SULFIDES + SO_2

Equilibrium Constant vs Temp.

① $3FeS + 2SO_2 \rightleftharpoons Fe_3O_4 + \frac{5}{2} S_2$
② $2ZnS + SO_2 \rightleftharpoons 2ZnO + \frac{3}{2} S_2$
③ $2PbS + SO_2 \rightleftharpoons 2PbO + \frac{3}{2} S_2$
④ $2Cu_2S + SO_2 \rightleftharpoons 2Cu_2O + \frac{3}{2} S_2$

+ CA 67,34508 V

EXHIBIT 6-23
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: H_2S CASE NO. -
PROCESS: High Temp in Situ H_2S GAS NO. 2A
Fixed Capital, Battery Limits Basis $2,320,000
Sulfur Production 167 NTD x 330 Days = 55,200 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	21.4 MCF	$.30 Avg.	$ 7.23	$ 398.0
1.	Total Raw Materials	-	-	$ 7.23	$ 398.0
2.	Direct Labor	18,000 MH	$3.50	1.15	63.0
3.	Supervision	1.	$10,000	.18	10.0
4.	Maintenance	5% FC	-	2.11	116.0
5.	Plant Supplies	10%(L2+L3)	-	.13	7.3
	Utilities:				
6. a)	Electricity	1,400 KWH	$.01	2.00	110.0
b)	Cooling Water	- MG	.05	-	-
c)	Boiler Feed Water	1.2 MG	.25	.30	16.5
d)	Fuel Gas	16.5 MCF	.30	4.95	272.0
7.	Total Utilities	-	-	7.25	398.5
8.	Other Direct (Catalyst)	50 NTY	$260	.24	13.0
9.	TOTAL DIRECT	-	-	$ 18.29	$ 1,005.8
10.	Payroll Burden	30% L2	-	.34	18.9
11.	Plant Overhead	70%(L2+L3)	-	.93	51.1
12.	Pack & Ship	1000 MH	3.90	.07	3.9
13.	Waste Disposal	1000 MH	3.90	.07	3.9
14.	Other Indirect	-	-	-	-
15.	TOTAL INDIRECT	-	-	$ 1.41	$ 77.8
16.	Depreciation	10% FC	-	4.22	232.0
17.	Taxes	2% FC	-	.84	46.4
18.	Insurance	1% FC	-	.42	23.2
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$ 5.48	$ 301.6
21.	TOTAL OPERATING	-	-	$ 25.18	$ 1,385.2
22.	Credits	None		-	-
23.	NET OPERATING	-	-	$ 25.18	$ 1,385.2

EXHIBIT 6-24

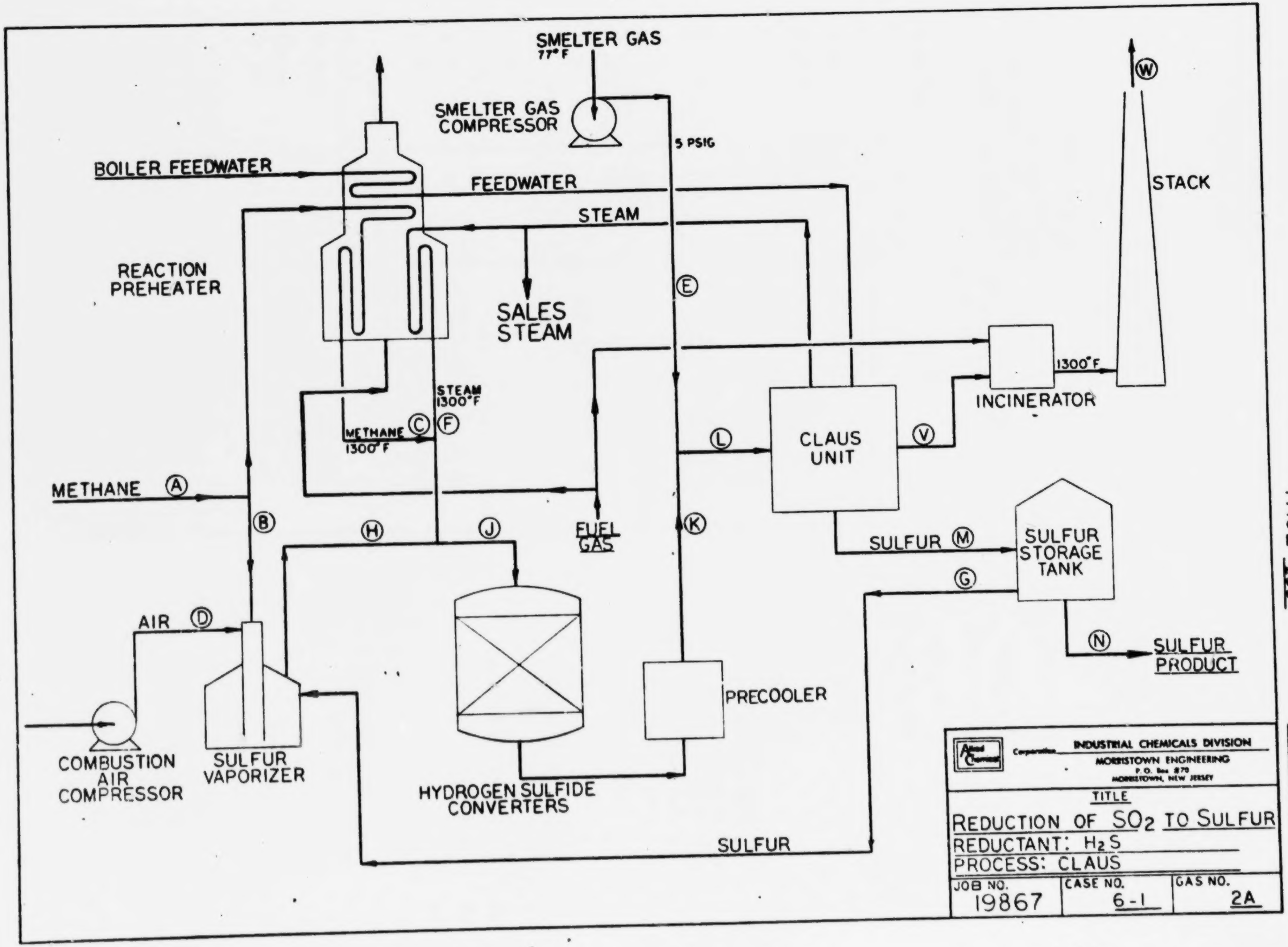
H_2S REDUCTION OF SO_2

CASE 6-1

SUMMATION OF MAJOR EQUIPMENT

1.	Combustion Air Compressor. - two stage centrifugal compressor to handle 9000 SCFM at an inlet temperature of 77°F with a P of 10 psi. Equipped with a 600 HP motor.	$ 60,000
2.	Sulfur Vaporizer - submerged combustion heater 53MM BTU/HR to vaporize 500 NT sulfur/day to exit at 1300°F burning fuel gas or methane at controlled excess air conditions. A small amount of SO_2 will be generated by this unit for later recovery.	$ 50,000
3.	Smelter Gas Compressor - single stage centrifugal compressor to handle 20,000 SCFM at an inlet temperature of 77°F with a P of 5 psi. Equipped with a 700 HP motor.	$ 60,000
4.	Preheater - vertical bottom fired process heater, radiant heat duty 7.7MM BTU/HR., convection heat duty 2MM BTU/HR., including economizer. Net heat release 14.5MM BTU/HR.	$ 60,000
5.	Hydrogen Sulfide Synthesis Convertors - two 13 foot diameter shop fabricated vertical vessels packed to a depth of 10 feet with alumina type catalyst.	$100,000
6.	Claus Unit - partial Claus sulfur recovery unit consisting of precooler, two stage converter, two sulfur condensers, coalescer, incinerator and exit stack. Unit is designed for a throughput of 37,000 SCFM with high temperature incineration and 150° exit stack in keeping with Allied Chemical's standard design. Incinerator takes 48MM BTU/HR. to burn residual H_2S.	$400,000
7.	Sulfur Storage Tank - vertical steel tank 2,000 ton capacity with steam coils and external insulation.	$ 40,000
	Total Purchased Equipment	$770,000
	Total Fixed Capital	$2,320,000

EXHIBIT 6-26


SMELTER GAS
77°F
SMELTER GAS COMPRESSOR
5 PSIG
BOILER FEEDWATER
FEEDWATER
STEAM
REACTION PREHEATER
SALES STEAM
STEAM 1300°F
METHANE 1300°F
Ⓒ
Ⓕ
Ⓔ
METHANE
Ⓐ
Ⓑ
Ⓗ
Ⓙ
FUEL GAS
Ⓚ
Ⓛ
CLAUS UNIT
Ⓥ
INCINERATOR
1300°F
STACK
Ⓦ
SULFUR
Ⓜ
SULFUR STORAGE TANK
Ⓖ
Ⓝ
SULFUR PRODUCT
AIR
Ⓓ
COMBUSTION AIR COMPRESSOR
SULFUR VAPORIZER
HYDROGEN SULFIDE CONVERTERS
PRECOOLER
SULFUR

Allied Chemical Corporation INDUSTRIAL CHEMICALS DIVISION
MORRISTOWN ENGINEERING
P. O. Box 870
MORRISTOWN, NEW JERSEY

TITLE
REDUCTION OF SO_2 TO SULFUR
REDUCTANT: H_2S
PROCESS: CLAUS

JOB NO.	CASE NO.	GAS NO.
19867	6-1	2A

DRAWING NO. 37181

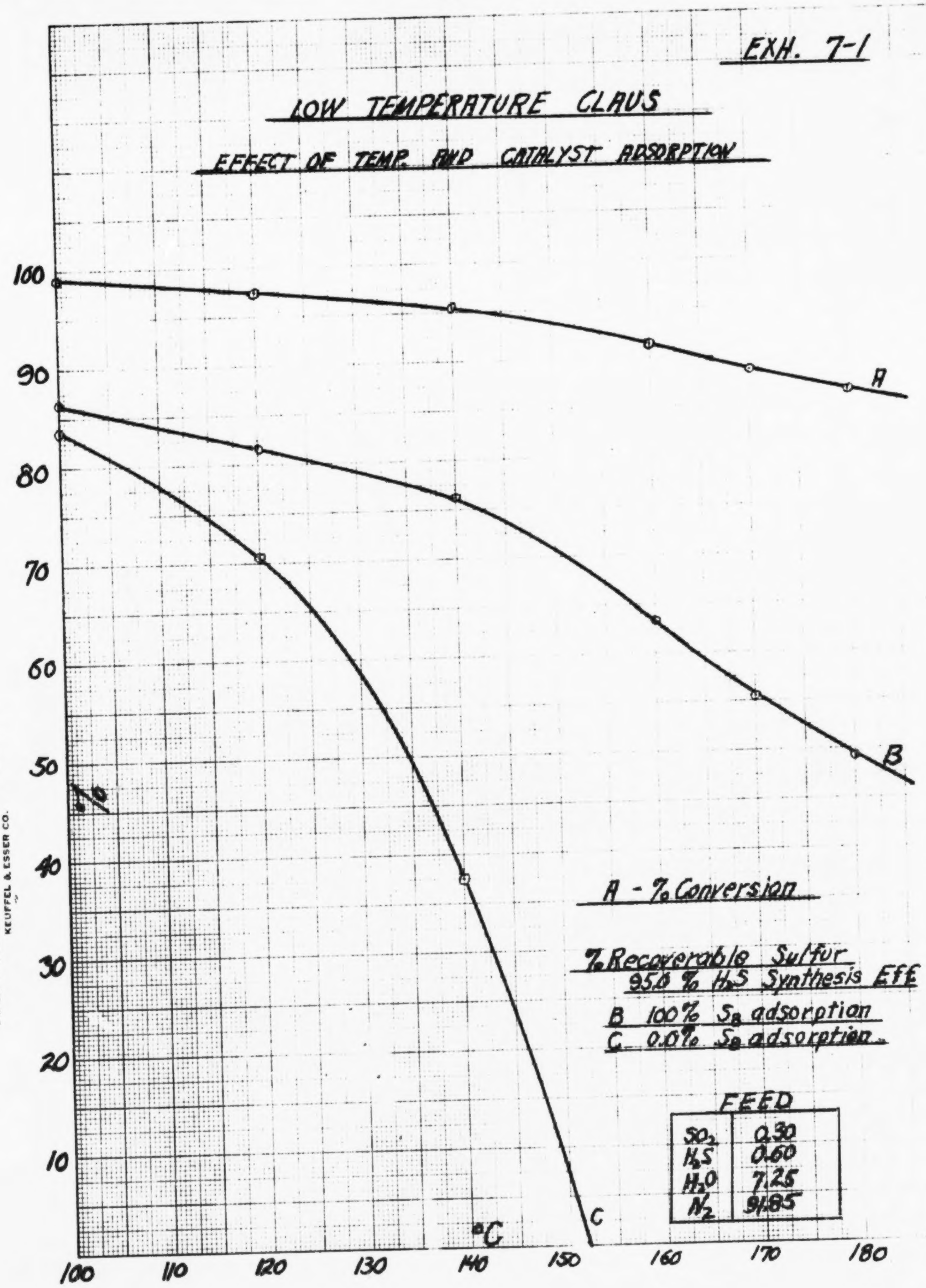
EXHIBIT 6-38
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: Coke CASE NO. 4
PROCESS: Lepsoe - Cominco at Trail GAS NO. 1

Fixed Capital, Battery Limits Basis $ 2,600,000
Sulfur Production 400 NTD x 330 Days = 132,000 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	MCF	$.30 Avg.	$ -	$ -
	Coke, 3" Furnace Grade	0.45	$24. Avg.	10.85	1,430.0
1.	Total Raw Materials	-	-	$ 10.85	$ 1,430.0
2.	Direct Labor	18,000 MH	$3.50	$ .475	63.0
3.	Supervision	1.	$10,000	.075	10.0
4.	Maintenance	5% FC	-	.985	130.0
5.	Plant Supplies	10%(L2+L3)	-	.055	7.3
	Utilities:				
6. a)	Electricity	80 KWH	$.01	.80	105.0
b)	Cooling Water 160 gpm	0.58 MG	.05	.03	4.0
c)	Boiler Feed Water	- MG	.25	-	-
d)	Fuel Gas	0.865 MCF	.30	.22	29.0
7.	Total Utilities	-	-	1.05	138.0
8.	Other Direct (Catalyst)	5 NTY	100/NT	-	0.5
9.	TOTAL DIRECT	-	-	$ 13.49	$ 1,778.8
10.	Payroll Burden	30% L2	-	.14	18.9
11.	Plant Overhead	70%(L2+L3)	-	.39	51.1
12.	Pack & Ship	1000 MH	3.90	.03	3.9
13.	Waste Disposal	1000 MH	3.90	.03	3.9
14.	Other Indirect	-	-	-	-
15.	TOTAL INDIRECT	-	-	$ .59	$ 77.8
16.	Depreciation	10% FC	-	1.97	260.0
17.	Taxes	2% FC	-	.39	52.0
18.	Insurance	1% FC	-	.20	26.0
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$ 2.56	$ 338.0
21.	TOTAL OPERATING	-	-	$ 16.64	$ 2,194.6
22.	Credits HP Steam	0.96 M Lbs.	.50	.48	63.2
23.	NET OPERATING	-	-	$ 16.16	$ 2,131.4



EXH. 7-1

LOW TEMPERATURE CLAUS

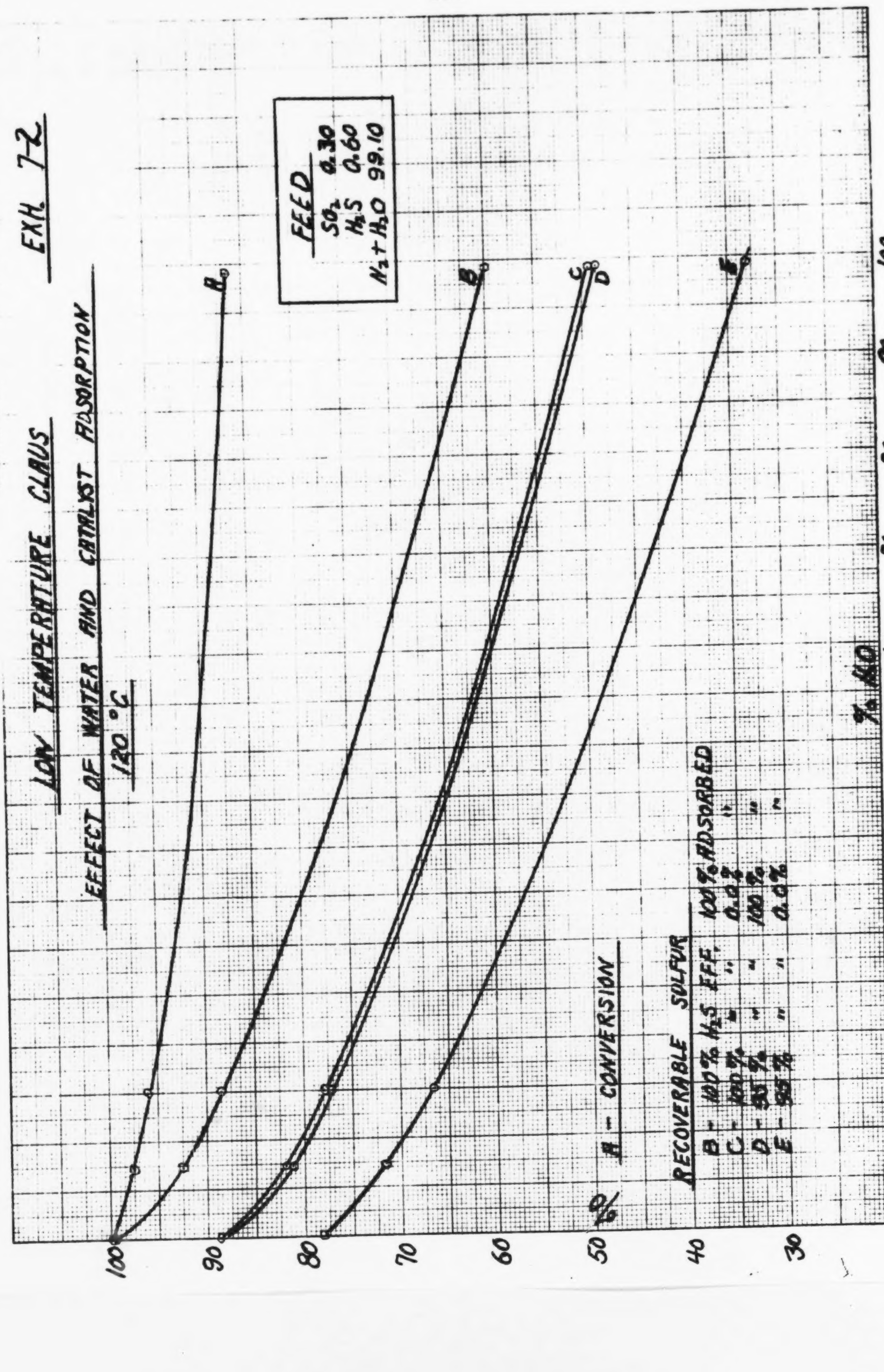
EFFECT OF TEMP. AND CATALYST ADSORPTION

FEED

SO_2	0.30
H_2S	0.60
H_2O	7.25
N_2	91.85

K&E 10 X 10 TO THE CENTIMETER 46 1510
18 X 25 CM MADE IN U.S.A.
KEUFFEL & ESSER CO.

EXHIBIT 7-2



K&E 10 X 10 TO THE CENTIMETER 46 1510
KEUFFEL & ESSER CO.

EXHIBIT 7-3
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: Coke — CASE NO. 14

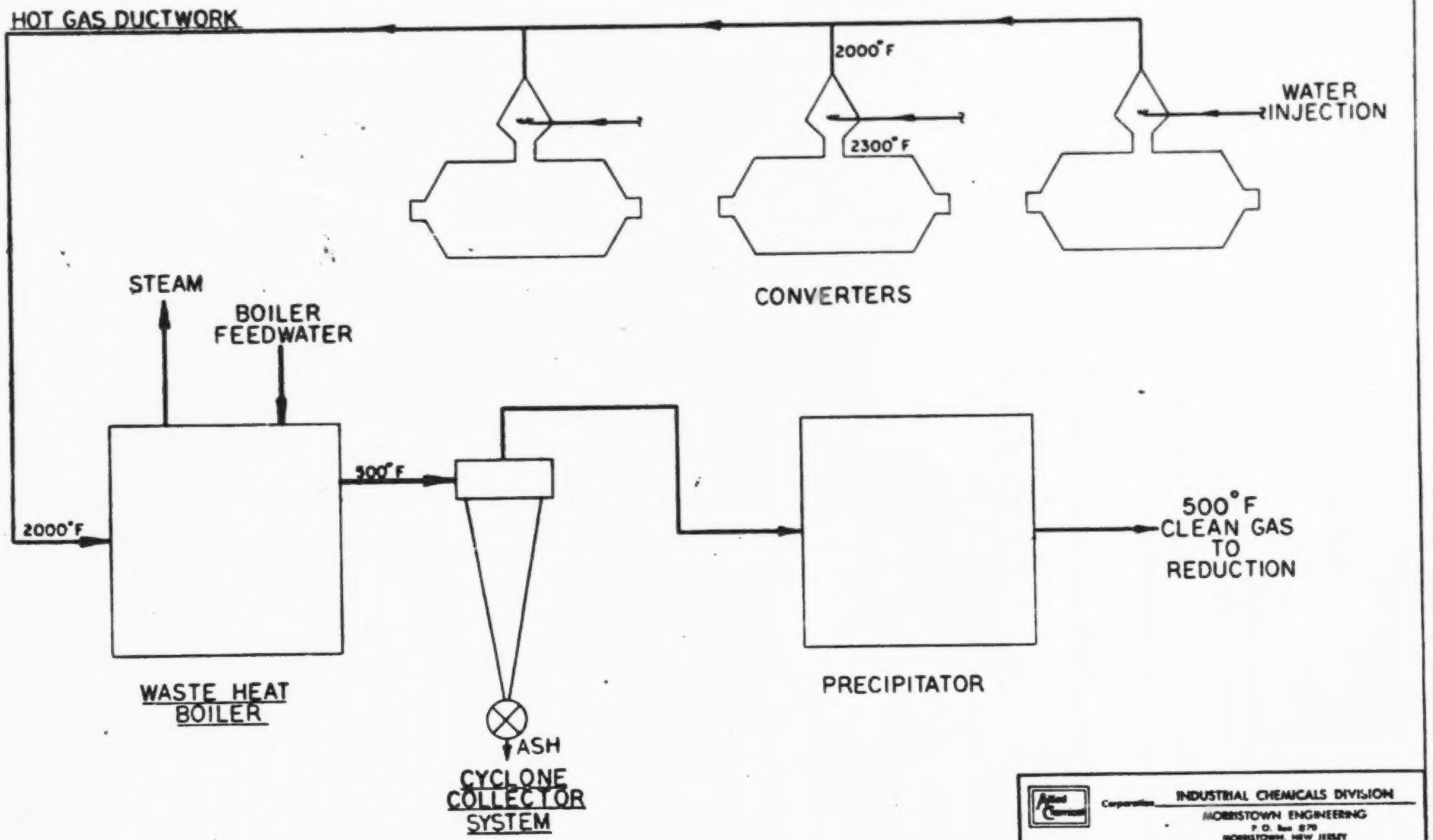
PROCESS: Topsoe - Heat Recovered — GAS NO. 2A

Fixed Capital, Battery Limits Basis — $ 4,740,000

Sulfur Production 195 NTD x 330 Days = 64,400 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	- MCF	$.30 Avg.	$ -	$ -
	Coke	0.516 NT	$24. Avg.	12.40	800.0
1.	Total Raw Materials	-	-	$ 12.40	$ 800.0
2.	Direct Labor	18,000 MH	$3.50	.98	63.0
3.	Supervision	1.	$10,000	.15	10.0
4.	Maintenance	5% FC	-	3.68	237.0
5.	Plant Supplies	10%(L2+L3)	-	.11	7.3
	Utilities:				
6. a)	Electricity 3,915 KW	483 KWH	$.01	4.83	311.0
b)	Cooling Water	2.8 MG	.05	.14	9.0
c)	Boiler Feed Water	0.65 MG	.25	.16	10.3
d)	Fuel Gas	4.3 MCF	.30	1.29	83.0
7.	Total Utilities	-	-	6.42	413.3
8.	Other Direct (Catalyst)	5 NTY	$100.	.01	0.5
9.	TOTAL DIRECT	-	-	$ 23.75	$ 1,531.1
10.	Payroll Burden	30% L2	-	.29	18.9
11.	Plant Overhead	70%(L2+L3)	-	.79	51.0
12.	Pack & Ship	1000 MH	3.90	.06	3.9
13.	Waste Disposal	1000 MH	3.90	.06	3.9
14.	Other Indirect	-	-	-	-
15.	TOTAL INDIRECT	-	-	$ 1.20	$ 77.7
16.	Depreciation	10% FC	-	7.35	474.0
17.	Taxes	2% FC	-	1.47	94.8
18.	Insurance	1% FC	-	.74	47.4
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$ 9.56	$ 616.2
21.	TOTAL OPERATING	-	-	$ 34.51	$ 2,225.0
22.	Credits HP Steam	4.92 M Lb.	.50	2.46	158.2
23.	NET OPERATING	-	-	$ 32.05	$ 2,066.8

EXHIBIT 7-4


HOT GAS DUCTWORK
2000°F
2300°F
WATER INJECTION
CONVERTERS
STEAM
BOILER FEEDWATER
2000°F
500°F
WASTE HEAT BOILER
ASH
CYCLONE COLLECTOR SYSTEM
PRECIPITATOR
500°F CLEAN GAS TO REDUCTION

Allied Chemical Corporation INDUSTRIAL CHEMICALS DIVISION
MORRISTOWN ENGINEERING
P.O. Box 270
MORRISTOWN, NEW JERSEY

TITLE
SMELTER GAS
COOLING

JOB NO. 19867
GAS NO. 2A

DRAWING NO. 37182

EXHIBIT 7-5

SMELTER GAS COOLING SYSTEM

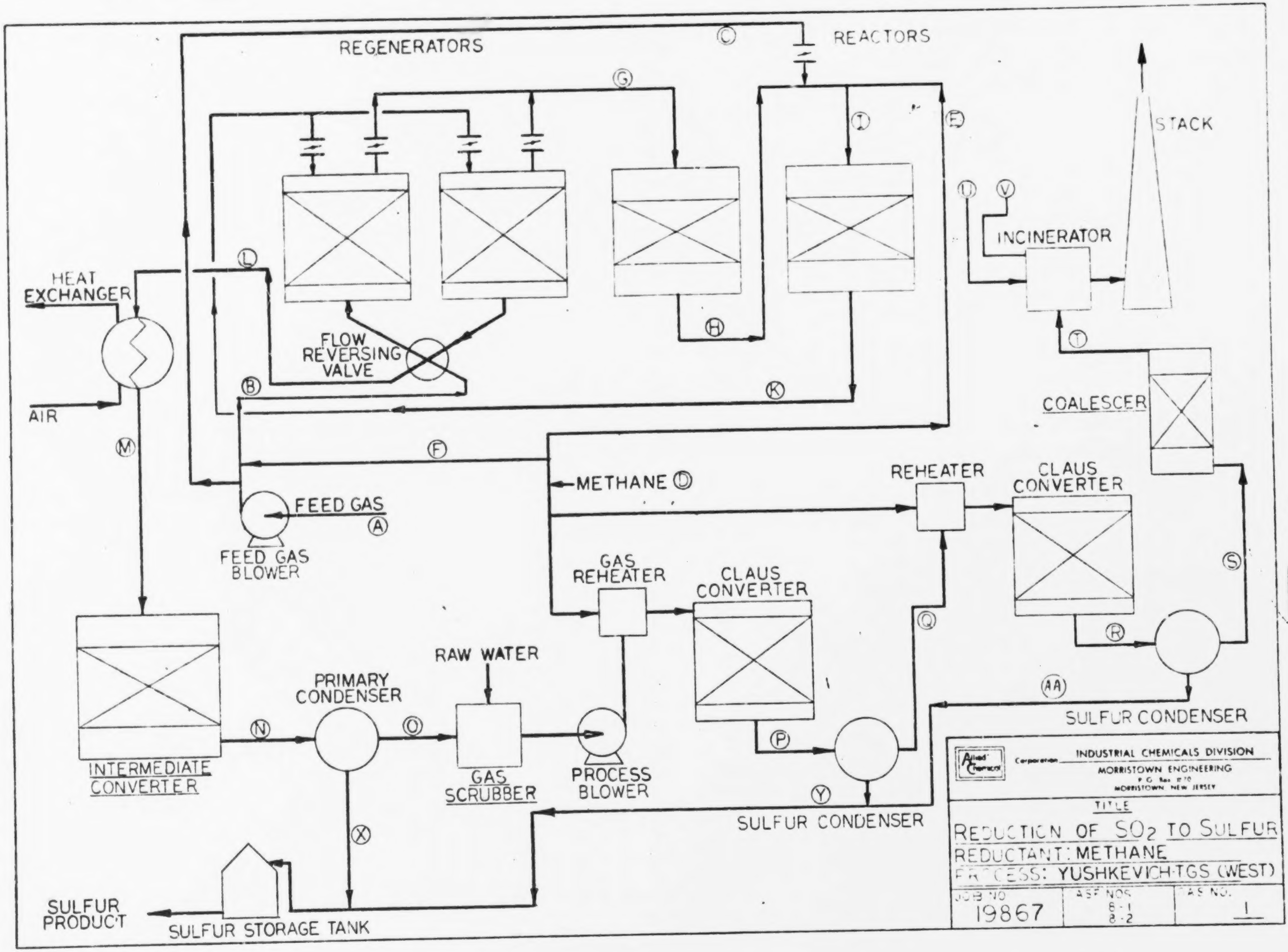
SUMMATION OF MAJOR EQUIPMENT

1.	Hot gas ductwork manifold - suitable for 21,000 SCFM 2000°F smelter gas, refractory lined steel.	$ 50,000
2.	Waste heat boiler - 40,000 lb./hr. at 450 psig water tube design, with soot blowers and auxiliaries.	$ 80,000
3.	Cyclone collector system - multiple steel cyclone collectors with ash gates.	$ 10,000
4.	Precipitator - plate and wire or rod curtain design electrostatic precipitator, carbon steel construction, for 21,000 SCFM 500°F gas.	$ 40,000
	Purchased Equipment Total	$180,000

CAPITAL COST ESTIMATE SUMMARY

	Item	Cost $	% of Total
1.	Purchased Equipment	$180,000	33.4
2.	Equipment Erection	42,000	7.8
3.	Foundations & Platforms	13,000	2.5
4.	Piping & Flues	61,000	11.2
5.	Instrumentation	6,000	1.1
6.	Insulation	16,000	3.0
7.	Electrical	5,000	1.0
8.	Buildings	11,000	2.0
9.	Utilities	53,000	9.9
10.	Physical Plant Cost	387,000	--
11.	Engineering & Construction	60,000	11.0
12.	Direct Plant Cost	447,000	--
13.	Contractors Fee	44,000	8.1
14.	Contingency	49,000	9.0
15.	Fixed Capital Cost	540,000	100.0

EXHIBIT 8-1


REGENERATORS
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REACTORS
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Ⓘ
Ⓔ
STACK
Ⓤ
Ⓥ
INCINERATOR
HEAT EXCHANGER
Ⓛ
FLOW REVERSING VALVE
Ⓗ
Ⓣ
AIR
Ⓑ
Ⓚ
COALESCER
Ⓜ
Ⓕ
METHANE Ⓓ
REHEATER
CLAUS CONVERTER
FEED GAS
Ⓐ
FEED GAS BLOWER
GAS REHEATER
CLAUS CONVERTER
Ⓢ
Ⓠ
Ⓡ
RAW WATER
PRIMARY CONDENSER
Ⓐ
SULFUR CONDENSER
Ⓝ
Ⓞ
Ⓟ
INTERMEDIATE CONVERTER
GAS SCRUBBER
PROCESS BLOWER
Ⓨ
SULFUR CONDENSER
Ⓧ
SULFUR PRODUCT
SULFUR STORAGE TANK

Allied Chemical Corporation INDUSTRIAL CHEMICALS DIVISION
MORRISTOWN ENGINEERING
P.O. Box 1070
MORRISTOWN, NEW JERSEY

TITLE
REDUCTION OF SO_2 TO SULFUR
REDUCTANT: METHANE
PROCESS: YUSHKEVICH-TGS (WEST)

JOB NO.	ASF NOS.	DAS NO.
19867	8-1 8-2	1

DRAWING No. 37183

EXHIBIT 8-2

METHANE REDUCTION OF SULFUR DIOXIDE

7130 SCFM

YUSHKEVICH PROCESS

CASE 1

GAS 1

Process Points	A	B	C	D	E	F	G	H	I	K	N	O	P	Q	R	S
Temp. °F	104	99	104	77	77	77	1292	1681	1315	1609	705	261	350	261	300	261
Vol. MSCFM	7.13	7.81	0.93	3.21	1.60	1.60	7.81	9.41	11.94	13.25	13.41	10.74	10.84	10.55	10.60	10.45
Vol. MSCFM at T°F	8.18	8.86	1.07	3.51	1.75	1.75	27.80	41.03	43.10	55.78	31.78	15.78	17.87	15.49	16.39	15.35
Vol. % CH_4		20.55		100.00	100.00		20.55		13.43							
% SO_2	90.00	71.51	90.00				71.51	25.22	26.87	6.64	2.92	3.64	1.83	1.89	0.93	0.94
% H_2O	10.00	7.94	10.00				7.94	40.68	32.84	41.77	48.23	60.22	63.24	64.96	66.56	67.52
% H_2S										10.90	4.93	6.15	2.55	2.62	0.70	0.71
% CO_2								17.05	13.43	23.01	23.01	28.73	28.48	29.26	29.12	29.54
% CO										0.92	0.91	1.14	1.12	1.15	1.15	1.17
% S_2								17.05	13.43	15.31	20.00	0.12	2.78	0.12	1.54	0.12
% H_2										1.16						
% COS										0.29						
% N_2																
% O_2																

POINT	SULFUR DISTRIBUTION		
X	From Intermed. Converter Cond.	343.38	NT/Day
Y	From No. 1 Claus Cond.	37.14	"
AA	From No. 2 Claus Cond.	19.40	"
	Total (Incl. Coalescer)	399.92	"

OVERALL CONVERSION 98.47 %

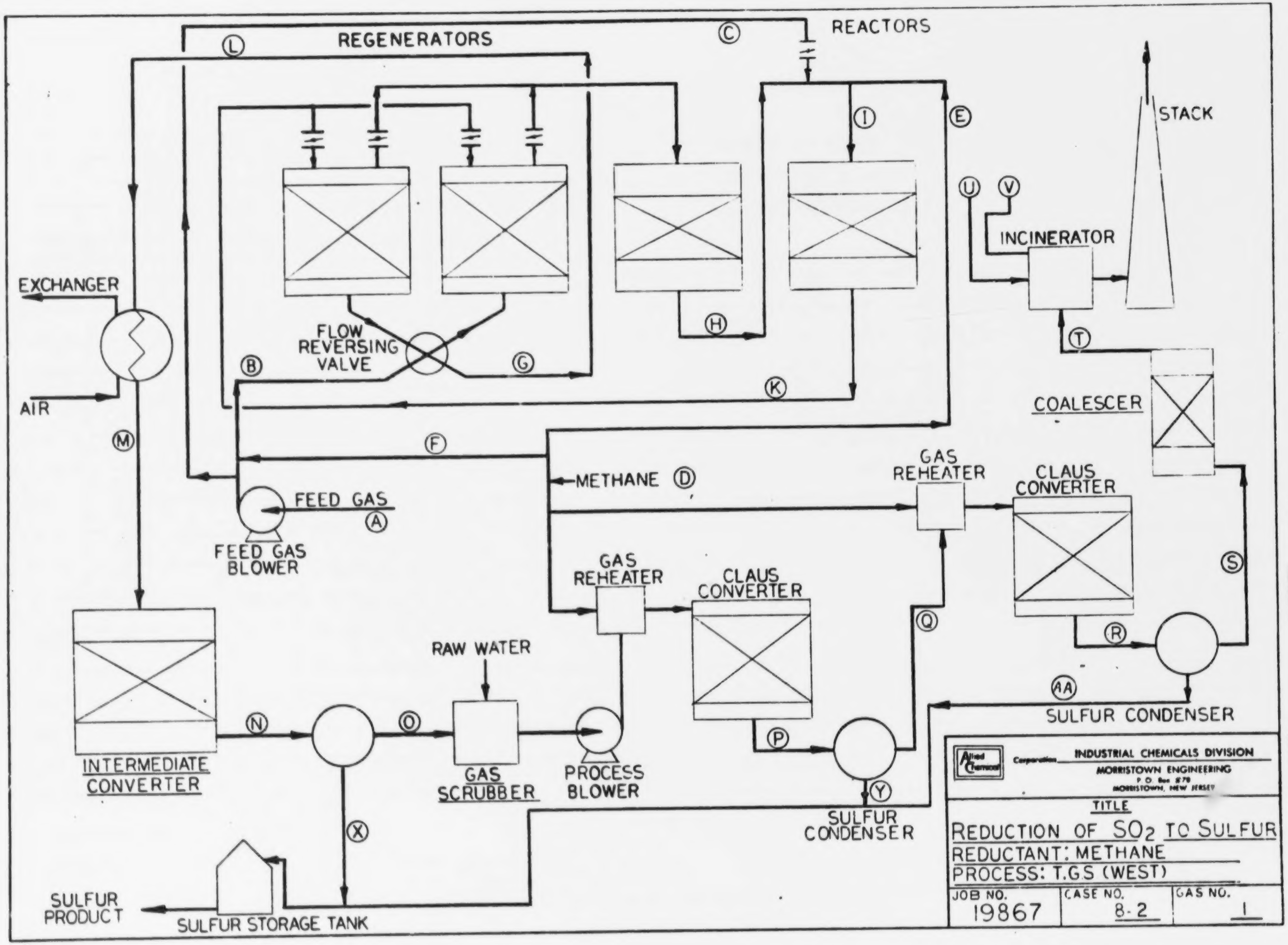
SULFUR YIELD 97.0 %

METHANE FOR RED'N: 11.6 MSCF/TON S

(12.2 at 60°F)

For Fuel: 0.72 MCF/Ton S

EXHIBIT 8-3


REGENERATORS
REACTORS
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Ⓘ
STACK
Ⓤ
Ⓥ
INCINERATOR
EXCHANGER
AIR
Ⓜ
FLOW REVERSING VALVE
Ⓑ
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Ⓚ
Ⓣ
COALESCER
Ⓕ
METHANE Ⓓ
FEED GAS Ⓐ
FEED GAS BLOWER
GAS REHEATER
CLAUS CONVERTER
GAS REHEATER
CLAUS CONVERTER
Ⓢ
Ⓠ
Ⓡ
RAW WATER
ⒶⒶ
SULFUR CONDENSER
INTERMEDIATE CONVERTER
Ⓝ
Ⓞ
GAS SCRUBBER
PROCESS BLOWER
Ⓟ
Ⓨ
SULFUR CONDENSER
Ⓧ
SULFUR PRODUCT
SULFUR STORAGE TANK

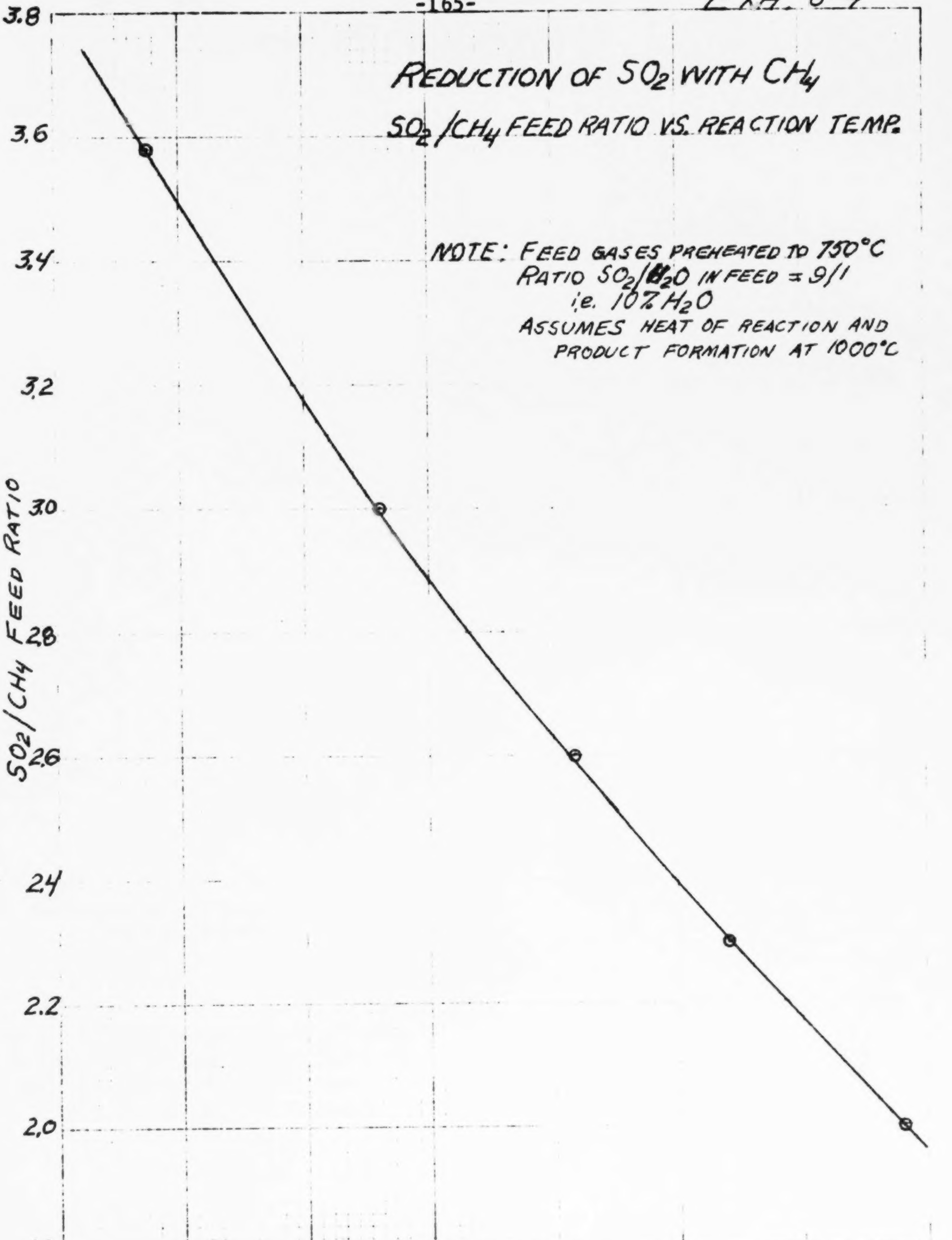
Allied Chemical Corporation INDUSTRIAL CHEMICALS DIVISION
MORRISTOWN ENGINEERING
P.O. Box 579
MORRISTOWN, NEW JERSEY

TITLE
REDUCTION OF SO_2 TO SULFUR
REDUCTANT: METHANE
PROCESS: T.G.S (WEST)

JOB NO.	CASE NO.	GAS NO.
19867	8-2	1

DRAWING NO. 37184

EXH. 8-4

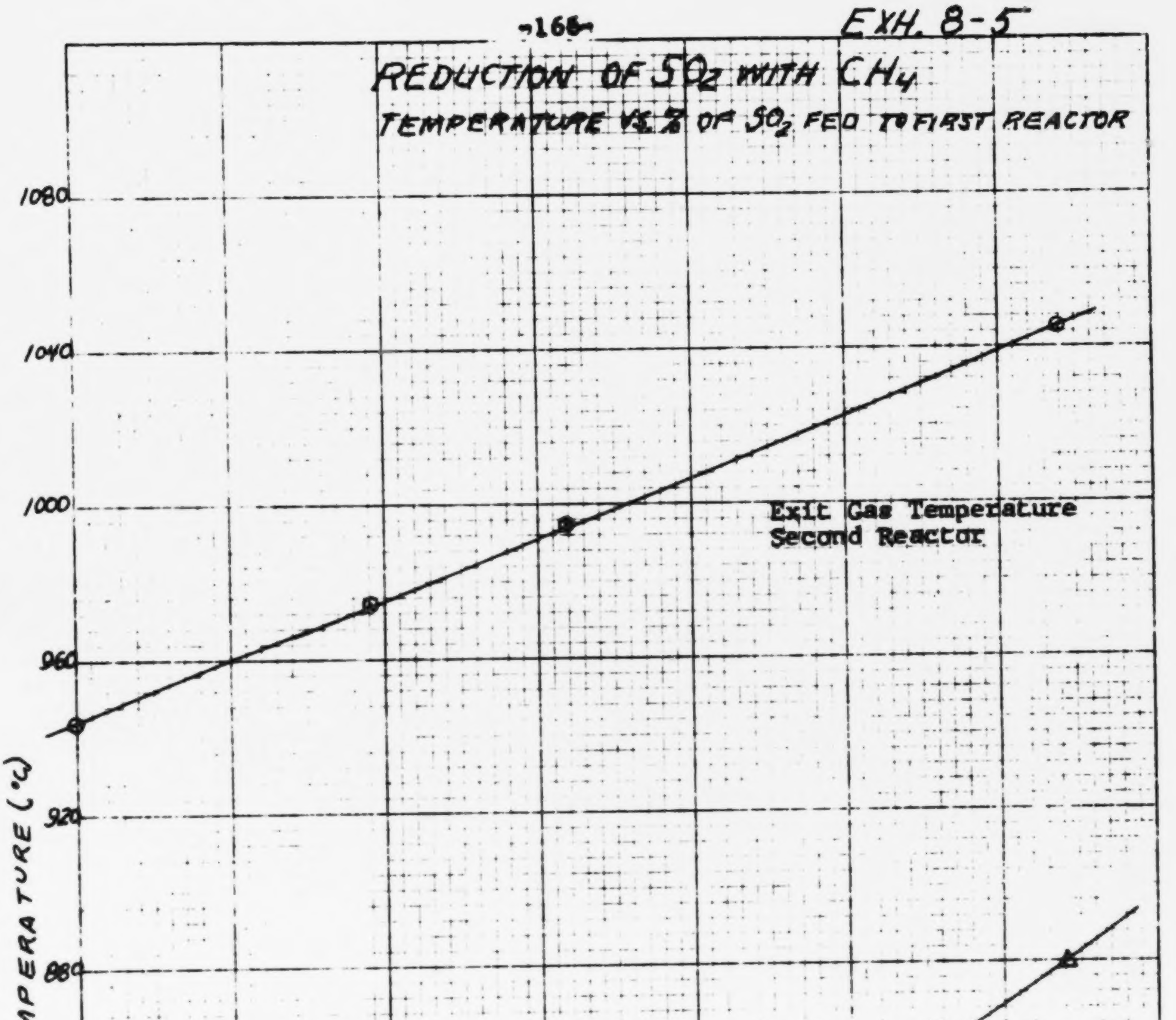


REDUCTION OF SO_2 WITH CH_4

SO_2/CH_4 FEED RATIO VS. REACTION TEMP.

NOTE: FEED GASES PREHEATED TO 750°C
RATIO SO_2/H_2O IN FEED = 9/1
i.e. 10% H_2O
ASSUMES HEAT OF REACTION AND
PRODUCT FORMATION AT 1000°C

EXH. 8-5



REDUCTION OF SO_2 WITH CH_4

TEMPERATURE VS. % OF SO_2 FED TO FIRST REACTOR

NOTE:
RATIO SO_2/CH_4 IN FEED TO 1ST REACTOR = 1.58
RATIO SO_2/H_2O IN FEED TO 1ST REACTOR = 9
ASSUMES HEAT OF REACTION AND PRODUCT FORMATION AT 1000°C

Exhibit 8-6

METHANE REDUCTION OF SULFUR DIOXIDE

CASE 2 | 7480 SCFM | WEST PROCESS

GAS 1

Process Points	A	B	C	D	E	F	G	H	I	K	N	O	P	Q	R	S
Temp. °F	77	77	77	77	77	77	1382	1747	1375	1636	750	261	350	261	300	261
Vol. MSCFM	7.48	8.42	0.75	3.37	1.68	1.69	8.42	10.05	10.80	13.98	13.81	11.66	11.91	11.18	11.26	11.02
Vol. MSCFM at T°F	8.18	9.21	0.82	3.68	1.84	1.85	31.58	45.12	40.28	59.55	31.21	17.13	19.62	16.37	17.41	16.13
Vol. % CH_4		20.09		100.00	100.00	100.00	20.09	0.00	13.42	0.00	0.00	0.00	0.00	0.00	0.00	0.00
% SO_2	90.00	71.92	90.00				71.92	29.13	28.86	6.70	5.85	6.92	2.71	2.89	1.43	1.47
% H_2O	10.00	7.99	10.00				7.99	36.00	29.61	41.95	42.47	50.26	57.38	61.10	63.54	64.93
% H_2S								3.63	2.92	9.75	11.69	13.84	5.42	5.78	2.87	2.93
% CO_2								16.20	13.05	22.23	24.38	28.86	28.27	30.11	29.89	30.55
% CO								0.53	0.43	1.58						
% S_2								13.72	11.06	15.72	15.61	0.12	6.22	12	2.27	0.12
% H_2								0.72	0.58	1.79						
% COS								0.07	0.07	0.28						
% N_2																
% O_2																

POINT	SULFUR DISTRIBUTION		
X	From Intermed. Converter Cond.	275.48	NT/Day
Y	From No. 1 Claus Cond.	93.43	"
AA	From No. 2 Claus Cond.	31.09	"
--	Total (Incl. Coalescer)	400.00	"

OVERALL CONVERSION 97.60 %

SULFUR YIELD 92.41 %

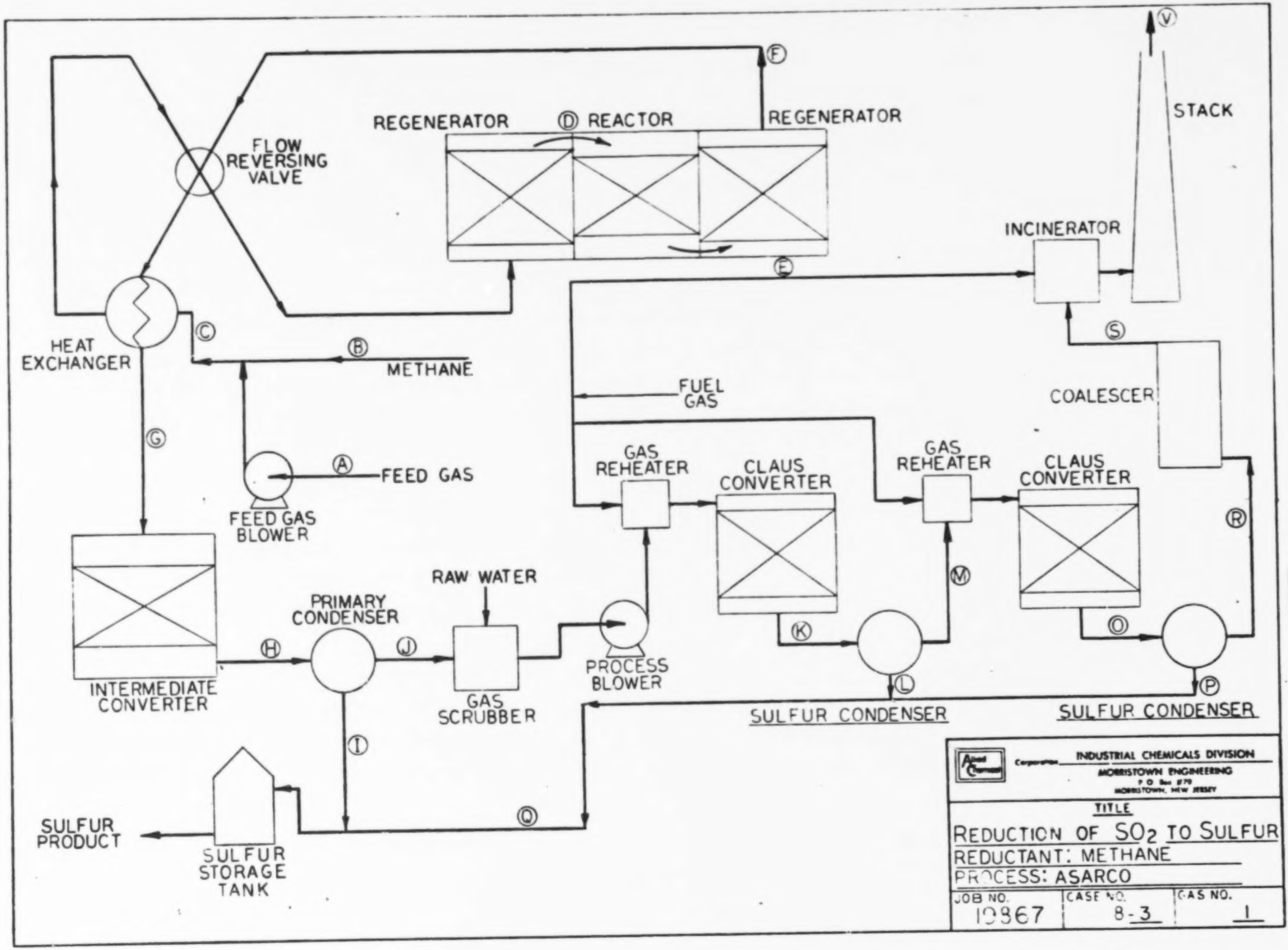
METHANE FOR RED'N: 12.0 MSCF/TON S

(12.7 at 60°F)

For Fuel: 0.72 MCF/Ton S

EXHIBIT 8-6

EXHIBIT 8-7


Ⓥ
STACK
Ⓕ
REGENERATOR
Ⓓ REACTOR
REGENERATOR
FLOW REVERSING VALVE
INCINERATOR
Ⓔ
HEAT EXCHANGER
Ⓒ
Ⓑ
METHANE
Ⓢ
FUEL GAS
COALESCER
Ⓖ
Ⓐ
FEED GAS
FEED GAS BLOWER
GAS REHEATER
CLAUS CONVERTER
GAS REHEATER
CLAUS CONVERTER
Ⓡ
RAW WATER
PRIMARY CONDENSER
Ⓜ
Ⓗ
Ⓙ
Ⓚ
Ⓞ
INTERMEDIATE CONVERTER
GAS SCRUBBER
PROCESS BLOWER
Ⓛ
Ⓟ
SULFUR CONDENSER
SULFUR CONDENSER
Ⓘ
Ⓠ
SULFUR PRODUCT
SULFUR STORAGE TANK

Allied Chemical Corporation INDUSTRIAL CHEMICALS DIVISION
MORRISTOWN ENGINEERING
P. O. Box 879
MORRISTOWN, NEW JERSEY

TITLE
REDUCTION OF SO_2 TO SULFUR
REDUCTANT: METHANE
PROCESS: ASARCO

JOB NO.	CASE NO.	GAS NO.
10867	8-3	1

DRAWING NO. 37185

Exhibit 8-7A

METHANE REDUCTION OF SULFUR DIOXIDE

7470 SCFM

ASARCO PROCESS

CASE 3

GAS 1

Process Points	A	B	C	D	E	F	G	H	J	K	M	O	R
Temp. °F	77	77	77	1992	2282	657	734	750	261	350	261	300	261
Vol. MSCFM	7.47	3.36	10.82	10.82	14.31	14.31	14.31	13.80	11.62	11.86	11.15	11.23	10.99
Vol. MSCFM at T°F	8.15	3.67	11.82	53.88	79.71	32.48	34.77	33.95	17.08	19.50	16.89	17.37	16.14
Vol. % CH_4		100.00	31.03	31.03									
% SO_2	90.00		62.07	62.07	8.27	8.27	8.27	5.67	6.73	2.64	2.81	1.39	1.42
% H_2O	10.00		6.90	6.90	41.24	41.24	41.24	42.78	50.78	57.69	61.33	63.69	65.05
% H_2S					6.29	6.29	6.29	11.34	13.46	5.28	5.61	2.79	2.85
% CO_2					17.87	17.87	17.87	24.36	28.91	28.34	30.13	29.92	30.56
% CO					5.36	5.36	5.36						
% S_2					16.08	16.08	16.08	15.85	0.12	6.05	0.12	2.21	0.12
% H_2					4.65	4.65	4.65						
% COS					0.24	0.24	0.24						
% N_2													
% O_2													

POINT	SULFUR DISTRIBUTION		
I	From Intermed. Converter Cond.	279.34	NT/Day
L	From No. 1 Claus Cond.	90.52	"
P	From No. 2 Claus Cond.	30.14	"
	Total (Incl. Coalescer)	400.00	"

OVERALL CONVERSION 97.67 %

SULFUR YIELD 92.63 %

METHANE FOR RED'N: 12.0 MSCF/TON S

(12.7 at 60°F)

For Fuel: 0.72 MCF/Ton S

Exhibit 8-8

METHANE REDUCTION OF SO_2

CASE 8-2 (WEST)

SUMMATION OF MAJOR EQUIPMENT

1. Feed Gas Flower - centrifugal gas blower to handle 7500 SCFM of gas at an inlet temperature of 77°F with a P of 60" W.C. equipped with a 200 HP motor.--------------------------------$12,000.

2. Reactor - Regenerator - carbon steel refractory brick lined chamber segmented into two regenerator sections and two reactor sections by means of refractory division walls. Each regenerator consists of 2000 cubic feet of packed refractory material on a refractory grillwork. Each reactor section consists of 1000 cubic feet alumina type catalyst supported on a refractory grillwork.---$250,000.

3. Flow Reversing Valve - similar to Allied present design, refractory lined four way piston valve.-----------------------$30,000.

4. Hot Gas Valves - four 2' diameter and one 1' diameter water cooled butterfly valves with refractory lined discs. These are specially designed valves developed by Allied Chemical and are included as major equipment.------------------------$35,000.

5. Heat Exchanger - shell and tuve process heat exchanger of type 309 S stainless steel construction.--------------------------$60,000.

6. Intermediate Converter - stainless steel conversion chamber approximately 16 ft. diameter by 18 ft. high complete with 1800 cubic feet of alumina catalyst supported on a cast iron grillwork, externally insulated.----------------------------------$70,000.

7. Primary Condenser - low pressure steam generator designed to condense sulfur from 14000 SCFM of 730°F gas at a rate of 280 short tons per day.---$60,000.

8. Gas Scrubber - spray chamber type gas cooler and condenser designed to cool 300°F gas to 150°F condensing excess moisture-$20,000.

9. Gas Reheaters (2) - natural gas fired auxiliary burners, each designed to reheat Claus process gas stream for feed to Claus Converters. Approximately reheater outlet temperature 45°F. Cost for two units.--$10,000.

10. Process Blower - two stage centrifugal pressure blower for 6 PSIG P equipped with 500 HP motor.----------------------$60,000.

11. Claus Converters (2) - units designed to convert H_2S - SO_2 to S at a 95% total conversion carbon steel construction with grillwork and refractory lining with a total of 1200 cubic feet of bauxite catalyst for the two units. Both units in common shell 12' diam. x 30' TL.---------------------------$50,000.

12. Sulfur Condensers (2) - shell and tube carbon steel condensers generating low pressure steam, complete with auxiliaries cost for two units.--$50,000.

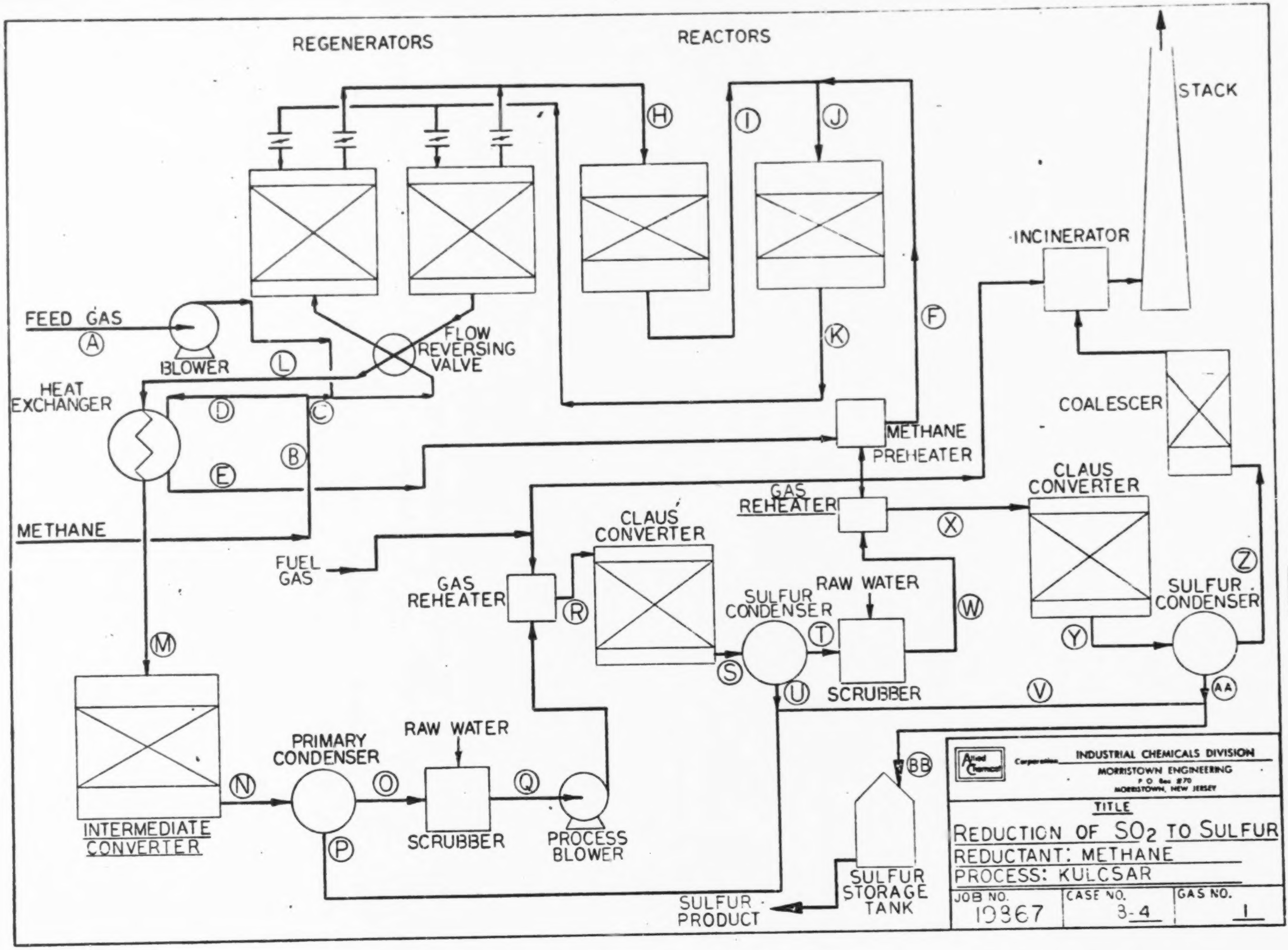
13. Coalescer - carbon steel circulating liquid sulfur type coalescer of standard design, complete with auxiliaries including pump.--$10,000.

14. Incinerator and stack - carbon steel refractory brick lined incinerator and 150 foot self supporting stack, to heat process effluent to 800°F and discharge.-----------------------$30,000.

15. Sulfur Storage Tank - vertical 5000 ton capacity with steam coils and external insulation.-----------------------------$70,000.

TOTAL Purchased Equipment-------------------------------$817,000.
TOTAL Fixed Capital------------------------------------$2,447,000

EXHIBIT 8-11


REGENERATORS
REACTORS
STACK
FEED GAS
(A)
BLOWER
(L)
FLOW REVERSING VALVE
HEAT EXCHANGER
(D)
(C)
(B)
(E)
(H)
(I)
(J)
(K)
(F)
INCINERATOR
COALESCER
METHANE PREHEATER
METHANE
FUEL GAS
GAS REHEATER
(R)
CLAUS CONVERTER
GAS REHEATER
(X)
CLAUS CONVERTER
(Z)
SULFUR CONDENSER
RAW WATER
(W)
(T)
(S)
(U)
SCRUBBER
(Y)
(V)
(AA)
(M)
INTERMEDIATE CONVERTER
(N)
PRIMARY CONDENSER
(O)
(P)
RAW WATER
SCRUBBER
(Q)
PROCESS BLOWER
SULFUR PRODUCT
(BB)
SULFUR STORAGE TANK
Allied Chemical Corporation INDUSTRIAL CHEMICALS DIVISION
MORRISTOWN ENGINEERING
P.O. Box 870
MORRISTOWN, NEW JERSEY
TITLE
REDUCTION OF SO2 TO SULFUR
REDUCTANT: METHANE
PROCESS: KULCSAR
JOB NO. 19367
CASE NO. 3-4
GAS NO. 1

DRAWING NO. 37186

Exhibit 8-11A

METHANE REDUCTION OF SULFUR DIOXIDE

CASE 4

GAS 1

6960 **SCFM**

KULCSAR **PROCESS**

Process Points	A	B	C	F	H	I	J	K	N	O	Q	S	T	W	Y	Z
Temp. °F	77	77	77	1695	1648	2012	1809	2012	842	275	130	550	275	115	500	275
Vol. MSCFM	6.96	3.13	2.09	1.04	9.05	11.10	12.14	13.12	12.85	10.83	6.26	6.61	5.56	3.59	3.61	3.57
Vol. MSCFM at T°F	7.62	3.42	2.28	1.14	38.82	55.83	56.09	65.86	34.05	16.20	7.52	13.57	8.31	4.35	7.05	5.34
Vol. % CH_4		100.00	100.00	100.00	23.10		8.57									
% SO_2	90.00				69.21	22.23	20.32	7.10	5.65	6.71	11.61	0.45	0.53	0.82	0.15	0.15
% H_2O	10.00				7.69	38.44	35.15	41.94	42.82	50.84	14.98	35.29	41.92	10.09	11.40	11.51
% H_2S						4.07	3.73	8.27	11.31	13.42	23.21	0.89	1.06	1.64	0.29	0.30
% CO_2						17.38	15.89	20.74	24.35	28.91	50.00	47.36	56.26	87.10	86.80	87.69
% CO						1.33	1.21	2.84								
% S_2						15.00	13.71	16.03	15.87	0.12	0.20	16.01	0.23	0.35	1.36	0.35
% H_2						1.43	1.31	2.81								
% COS						0.12	.11	0.27								
% N_2																
% O_2																

POINT	SULFUR DISTRIBUTION		
P	From Intermed. Converter Cond.	260.75	NT/Day
U	From No. 1 Claus Cond.	134.51	"
AA	From No. 2 Claus Cond.	4.68	"
BB	Total (Incl. Coalescer)	399.84	"

OVERALL CONVERSION 99.92 %

SULFUR YIELD 99.32 %

METHANE FOR RED'N: 11.3 MSCF/TON S

(11.9 at 60°F)

For Fuel: 1.0 MCF/Ton S

Exhibit 8-12

METHANE REDUCTION OF SO_2

CASE 8-1 (YUSHKEVICH)

SUMMATION OF MAJOR EQUIPMENT

1. Feed Gas Blower - centrifugal gas blower to handle 7500 SCFM of gas at an inlet temperature of 77°F. with a P of 60" W.C. equipped with a 200 HP motor.--------------------------------$12,000.

2. Reactor - Regenerator - carbon steel refractory brick lined chamber segmented into two regenerator sections and two reactor sections by means of refractory division walls. Each regenerator consists of 2000 cubic feet of packed refractory material on a refractory grillwork. Each reactor section consists of 1450 cubic feet alumina type catalyst supported on a refractory grillwork.--$280,000.

3. Flow Reversing Valve - similar to Allied present design, refractory lined four way piston valve.-----------------------$30,000.

4. Hot Gas Valves - four 2' diameter and one 1' diameter water cooled butterfly valves with refractory lined discs. These are specially designed valves developed by Allied Chemical and are included as major equipment.-----------------------$35,000.

5. Heat Exchanger - shell and tube process heat exchanger of type 309 S stainless steel construction.------------------------$60,000.

6. Intermediate Converter - stainless steel conversion chamber approximately 16 ft. diameter by 18 ft. high complete with 1800 cubic feet of alumina catalyst supported on a cast iron grillwork, externally insulated.--------------------------------$70,000.

7. Primary Condenser - low pressure steam generator designed to condense sulfur from 14000 SCFM of 730°F. gas at a rate of 280 short tons per day.---$60,000.

8. Gas Scrubber - spray chamber type gas cooler and condenser designed to cool 300°F. gas to 150°F. condensing excess moisture. --$20,000.

9. Gas Reheaters (2) - natural gas fired auxiliary burners, each designed to reheat Claus process gas stream for feed to Claus Converters. Approximately reheater outlet temperature 450°F. Cost for two units.--$10,000.

10. Process Blower - two stage centrifugal pressure blower for 6 PSIG P equipped with 500 HP motor.-----------------------$60,000.

11. Claus Converters (2) - units designed to convert H_2S - SO_2 to S at a 95% total conversion carbon steel construction with grillwork and refractory lining with a total of 1200 cubic feet of bauxite catalyst for the two units. Both units in common shell 12' diam. x 30' TL.----------------------------$50,000.

12. Sulfur Condensers (2) - shell and tube carbon steel condensers generating low pressure steam, complete with auxiliaries, cost for two units.---$50,000.

13.. Coalescer - carbon steel circulating liquid sulfur type coalescer of standard design, complete with auxiliaries including pump.---$10,000.

14. Incinerator and stack - carbon steel refractory brick lined incinerator and 150 foot self supporting stack, to heat process effluent to 800°F. and discharge.----------------------$30,000.

15. Sulfur Storage Tank - vertical 5000 ton capacity with steam coils and external insulation.------------------------------$70,000.

TOTAL Purchased Equipment ---------------------------------$847,000.
TOTAL Fixed Capital-------------------------------------$2,540,000.

Exhibit 8-14

METHANE REDUCTION OF SO_2

CASE 8-3 (ASARCO)

SUMMATION OF MAJOR EQUIPMENT

1. Feed Gas Blower - centrifugal gas blower to handle 7500 SCFM of gas at an inlet temperature of 77°F with a P of 50" W.C. equipped with a 150 HP motor.------------------------------$10,000.

2. Reactor-Regenerator - carbon steel refractory brick lined chamber segmented into two regenerator sections and one reactor section by means of refractory division walls. Each regenerator consists of 2000 cubic feet of packed refractory material on a refractory grillwork. Reactor section consists of an 8 foot deep Kruzite refractory checkerwork with approximately 50% void space, supported on a similar refractory grillwork.--$170,000.

3. Flow Reversing Valve - similar to Allied present design, refractory lined four way piston valve.----------------------$30,000.

4. Heat Exchanger - shell and tube process heat exchanger of type 309S stainless steel construction.---------------------$60,000.

5. Intermediate Converter - stainless steel conversion chamber approximately 16 ft. diameter by 18 ft. high complete with 1800 cubic feet of alumina catalyst supported on a cast iron grillwork, externally insulated.----------------------------$70,000.

6. Primary Condenser - low pressure steam generator designed to condense sulfur from 14000 SCFM of 730°F gas at a rate of 280 short tons per day.--------------------------------------$60,000.

7. Gas Scrubber - spray chamber type gas cooler and condenser designed to cool 300°F gas to 150°F condensing excess moisture.--$20,000.

EXH 8-14 (Cont.)

8. Gas Reheaters (2) - Natural Gas fired auxiliary burners, each designed to reheat Claus process gas stream for feed to Claus Converters. Approximate reheater outlet temperature 450°F. Cost for two units.--$10,000.

9. Process Blower - two stage' centrifugal pressure blower for 6 PSIG P equipped with 500 HP motor.------------$60,000.

10. Claus Converters (2) - units designed to convert H_2S - SO_2 to S at a 95% total conversion carbon steel construction cubic feet of bauxite catalyst for the two units. Both units in common shell 12' diam. x 30' TL.------------------$50,000.

11. Sulfur Condensers (2) - shell and tube carbon steel condensers generating low pressure steam, complete with auxiliaries, cost for two units.--$50,000.

12. Coalescer - carbon steel circulating liquid sulfur type coalescer of standard design, complete with auxiliaries including pump.---$10,000.

13. Incinerator and Stack - carbon steel refractory brick lined incinerator and 150 foot self supporting stack, to heat process effluent to 800°F and discharge.-------------------$30,000.

14. Sulfur Storage Tank - vertical 5000 ton capacity with steam coils and external insulation.------------------------------$70,000.

TOTAL Purchased Equipment---------------------------------$700,000.
TOTAL Fixed Capital-------------------------------------$2,100,000.

Exhibit 8-15

METHANE REDUCTION OF SO_2

CASE 8-4 (HULCSAT)

SUMMATION OF MAJOR EQUIPMENT

1. Feed Gas Blower - centrifugal gas blower to handle 7500 SCFM of gas at an inlet temperature of 77°F with a P of 60" W.C. equipped with a 200 HP motor.-------------------------------$12,000.
2. Reactor - Regenerator - carbon steel refractory brick lined chamber segmented into two regenerator sections and two reactor sections by means of refractory division walls. Each regenerator consists fo 2000 cubic feet of packed refractory material on a refractory grillwork. First reactor section consists of 200 cubic feet silica gel type catalyst. Second Reactor consists of 1600 cubic feet silica gel catalyst. Including support grillwork.---$260,000.
3. Flow Reversing Valve - similar to Allied present design, refractory lined four way piston valve.----------------------$30,000.
4. Hot Gas Valves - four 2' diameter and one 1' diameter water cooled butterfly valves with refractory lined discs. These are specially designed valves developed by Allied Chemical and are included as major equipment.------------------------$35,000.
5. Heat Exchanger - shell and tube process heat exchanger of alonized steel construction.---------------------------------$20,000.
6. Intermediate Converter - stainless steelconversion chamber approximately 16 ft. diameter by 18 ft. high complete with 1800 cubic feet of alumina catalyst supported on a cast iron grillwork, externally insulated.----------------------------$70,000.
7. Primary Condenser - low pressure steam generator designed to condense sulfur from 1400 SCFM of 730 Gas at a rate of 280 short tons per day.--$60,000.

8. Gas Scrubbers (2) - spray chamber type gas cooler and condenser designed to cool 300°F to 115-130°F condensing excess moisture.------------------------------$40,000.

9. Gas Reheaters (2) - natural gas fired auxiliary burners, each designed to reheat Claus process gas stream for feed to Claus Converters. Approximately reheater outlet temperature 450°F. Cost for two units.------------------------------$12,000.

10. Methane preheater - natural gas direct fired auxiliary burner, designed to preheat methane to 1700°F.------------------------------$8,000.

11. Process Blower - two stage centrifugal pressure blower for 6 PSIG P equipped with 500 HP motor.------------------------------$60,000.

12. Claus Converters (2) - units designed to convert H_2S - SO_2 to S at a 95% total conversion carbon steel construction with grillwork and refractory lining with total of 1200 cubic feet of bauxite catalyst for the two units. Both units in common shell 12' diameter x 30' TL.------------------------------$50,000.

13. Sulfur Condensers (2) - shell and tube carbon steel condensers generating low pressure steam, complete with auxiliaries cost for two units.------------------------------$50,000.

14. Coalescer - carbon steel circulating liquid sulfur type coalescer of standard design, complete with auxiliaries including pump.------------------------------$10,000.

15. Incinerator and stack - carbon steel refractory brick lined incinerator and 150 foot self supporting stack, to heat process effluent to 800°F and discharge.------------------------------$30,000.

16. Sulfur Storage Tank - vertical 5000 ton capacity with steam coils and external insulation.------------------------------$70,000.

TOTAL Purchased Equipment------------------------------$817,000.
TOTAL Fixed Capital------------------------------$2,447,000.

EXHIBIT 8-20
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: Methane CASE NO. 8-1
PROCESS: Yushkevitch GAS NO. 1

Fixed Capital, Battery Limits Basis $2,540,000
Sulfur Production 388 NTD x 330 Days = 128,000 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	12.7 MCF	$.30 Avg.	$ 3.81	$ 488.0
1.	Total Raw Materials	-	-	$ 3.81	$ 488.0
2.	Direct Labor	18,000 MH	$3.50	.49	63.0
3.	Supervision	1.	$10,000	.08	10.0
4.	Maintenance	5% FC	-	.99	127.0
5.	Plant Supplies	10%(L2+L3)	-	.06	7.3
	Utilities:				
6. a)	Electricity	40 KWH	$.01	.40	51.2
b)	Cooling Water	1.8 MG	.05	.09	11.5
c)	Boiler Feed Water	- MG	.25	-	-
d)	Fuel Gas	0.72 MCF	.30	.22	28.2
7.	Total Utilities	-	-	.71	90.9
8.	Other Direct (Catalyst)	3.8 Lb.	.13	.50	64.0
9.	TOTAL DIRECT	-	-	$ 6.64	$ 850.2
10.	Payroll Burden	30% L2	-	.15	18.9
11.	Plant Overhead	70%(L2+L3)	-	.40	51.1
12.	Pack & Ship	1000 MH	3.90	.03	3.9
13.	Waste Disposal	1000 MH	3.90	.03	3.9
14.	Other Indirect	-	-	-	-
15.	TOTAL INDIRECT	-	-	$ .61	$ 77.8
16.	Depreciation	10% FC	-	1.98	254.0
17.	Taxes	2% FC	-	.40	50.8
18.	Insurance	1% FC	-	.20	25.4
19.	Other Fixed	-	-		
20.	TOTAL FIXED	-	-	$ 2.58	$ 330.2
21.	TOTAL OPERATING	-	-	$ 9.83	$1,258.2
22.	Credits L.P. Steam	1.6 M Lb.	.25	.40	51.2
23.	NET OPERATING	-	-	$ 9.43	$1,207.0

EXHIBIT 8-21
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: Methane　　CASE NO. 8-2
PROCESS: West - Texas Gulf Sulfur　　GAS NO. 1

Fixed Capital, Battery Limits Basis　　$2,447,000
Sulfur Production 370 NTD x 330 Days = 122,000 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	12.7 MCF	$.30 Avg.	$ 3.81	$ 465.0
1.	Total Raw Materials	-	-	$ 3.81	$ 465.0
2.	Direct Labor	18,000 MH	$3.50	.52	63.0
3.	Supervision	1.	$10,000	.09	10.0
4.	Maintenance	5% FC	-	1.00	122.4
5.	Plant Supplies	10%(L2+L3)	-	.06	7.3
	Utilities:				
6. a)	Electricity	40 KWH	$.01	.40	48.7
b)	Cooling Water	1.8 MG	.05	.09	11.0
c)	Boiler Feed Water	- MG	.25	-	-
d)	Fuel Gas	0.72 MCF	.30	.22	26.9
7.	Total Utilities	-	-	.71	86.6
8.	Other Direct (Catalyst)	3.8 Lb.	.13	.50	61.0
9.	TOTAL DIRECT	-	-	$ 6.69	$ 815.3
10.	Payroll Burden	30% L2	-	.15	18.9
11.	Plant Overhead	70%(L2+L3)	-	.42	51.1
12.	Pack & Ship	1000 MH	3.90	.03	3.9
13.	Waste Disposal	1000 MH	3.90	.03	3.9
14.	Other Indirect	-	-	-	-
15.	TOTAL INDIRECT	-	-	$ .63	$ 77.8
16.	Depreciation	10% FC	-	2.00	244.7
17.	Taxes	2% FC	-	.40	49.0
18.	Insurance	1% FC	-	.20	24.5
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$ 2.60	$ 318.2
21.	TOTAL OPERATING	-	-	$ 9.92	$1,211.3
22.	Credits L P Steam	1.6 M Lb.	.25	.40	48.8
23.	NET OPERATING	-	-	$ 9.52	$1,162.5

EXHIBIT 8-22a
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: Methane CASE NO. 8-3
PROCESS: Asarco GAS NO. 1

Fixed Capital, Battery Limits Basis $2,100,000
Sulfur Production 370 NTD x 330 Days = 122,000 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	12.7 MCF	$.30 Avg.	$ 3.81	$ 465.0
1.	Total Raw Materials	-	-	$ 3.81	$ 465.0
2.	Direct Labor	18,000 MH	$3.50	.52	63.0
3.	Supervision	1.	$10,000	.09	10.0
4.	Maintenance	5% FC	-	.86	105.0
5.	Plant Supplies	10%(L2+L3)	-	.06	7.3
	Utilities:				
6. a)	Electricity	38.4 KWH	$.01	.38	46.6
b)	Cooling Water	1.8 MG	.05	.09	11.0
c)	Boiler Feed Water	- MG	.25	-	-
d)	Fuel Gas	0.72 MCF	.30	.22	26.9
7.	Total Utilities	-	-	.69	84.5
8.	Other Direct (Catalyst)	0.76 Lb.	.13	.10	12.2
9.	TOTAL DIRECT	-	-	$ 6.13	$ 747.0
10.	Payroll Burden	30% L2	-	.15	18.9
11.	Plant Overhead	70%(L2+L3)	-	.42	51.1
12.	Pack & Ship	1000 MH	3.90	.03	3.9
13.	Waste Disposal	1000 MH	3.90	.03	3.9
14.	Other Indirect	-	-	-	-
15.	TOTAL INDIRECT	-	-	$ .63	$ 77.8
16.	Depreciation	10% FC	-	1.72	210.0
17.	Taxes	2% FC	-	.34	42.0
18.	Insurance	1% FC	-	.17	21.0
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$ 2.23	$ 273.0
21.	TOTAL OPERATING	-	-	$ 8.99	$ 1,097.8
22.	Credits L P Steam	1.6 M Lb.	.25	.40	48.8
23.	NET OPERATING	-	-	$ 8.59	$ 1,049.0

EXHIBIT 8-22b
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: Methane | CASE NO. 8-3 (Factored)
PROCESS: Asarco | GAS NO. 1

Fixed Capital, Battery Limits Basis $1,400,000
Sulfur Production 185 NTD x 330 Days = 61,000 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	12.7 MCF	$.30 Avg.	$ 3.81	$ 232.5
1.	Total Raw Materials	-	-	$ 3.81	$ 232.5
2.	Direct Labor	18,000 MH	$3.50	1.04	63.0
3.	Supervision	1.	$10,000	.17	10.0
4.	Maintenance	5% FC	-	1.15	70.0
5.	Plant Supplies	10%(L2+L3)	-	.12	7.3
	Utilities:				
6. a)	Electricity	38.4 KWH	$.01	.38	23.3
b)	Cooling Water	1.8 MG	.05	.09	5.5
c)	Boiler Feed Water	- MG	.25	-	-
d)	Fuel Gas	0.72 MCF	.30	.22	13.5
7.	Total Utilities	-	-	.69	42.3
8.	Other Direct (Catalyst)	0.76 Lb.	.13	.10	6.1
9.	TOTAL DIRECT	-	-	$ 7.08	$ 431.2
10.	Payroll Burden	30% L2	-	.31	18.9
11.	Plant Overhead	70%(L2+L3)	-	.84	51.1
12.	Pack' & Ship	1000 MH	3.90	.06	3.9
13.	Waste Disposal	1000 MH	3.90	.06	3.9
14.	Other Indirect	-	-	-	-
15.	TOTAL INDIRECT	-	-	$ 1.27	$ 77.8
16.	Depreciation	10% FC	-	2.30	140.0
17.	Taxes	2% FC	-	.46	28.0
18.	Insurance	1% FC	-	.23	14.0
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$ 2.99	$ 182.0
21.	TOTAL OPERATING	-	-	$11.34	$ 691.0
22.	Credits L.P. Steam	1.6 M Lb.	.25	.40	24.4
23.	NET OPERATING	-	-	$10.94	$ 666.6

EXHIBIT 8-23
PH 22-68-24

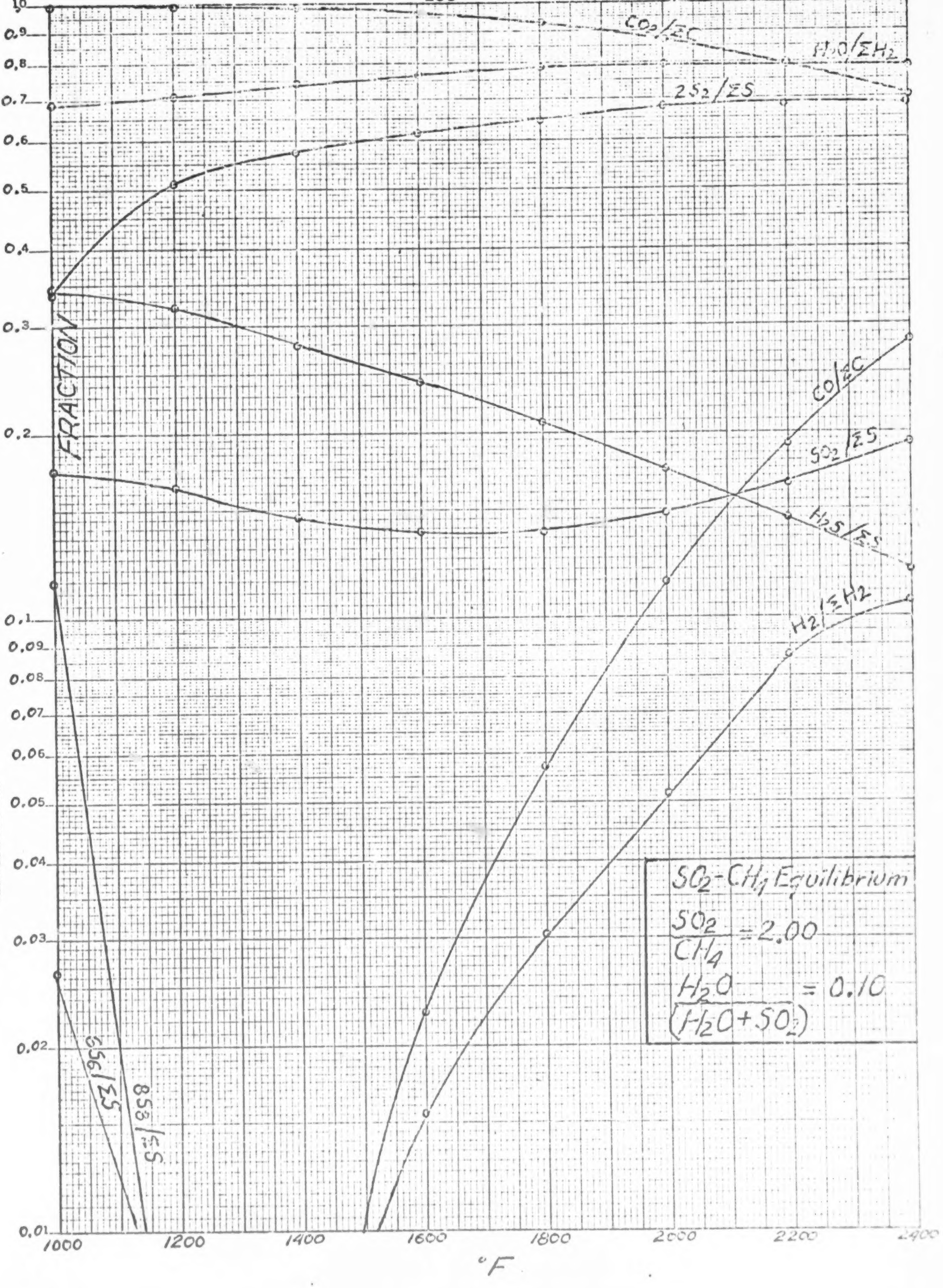
REDUCTION OF SO_2

REDUCTANT: Methane CASE NO. 8-4
PROCESS: Kulcsar GAS NO. 1

Fixed Capital, Battery Limits Basis $2,447,000
Sulfur Production 400 NTD x 330 Days = 132,000 NTY

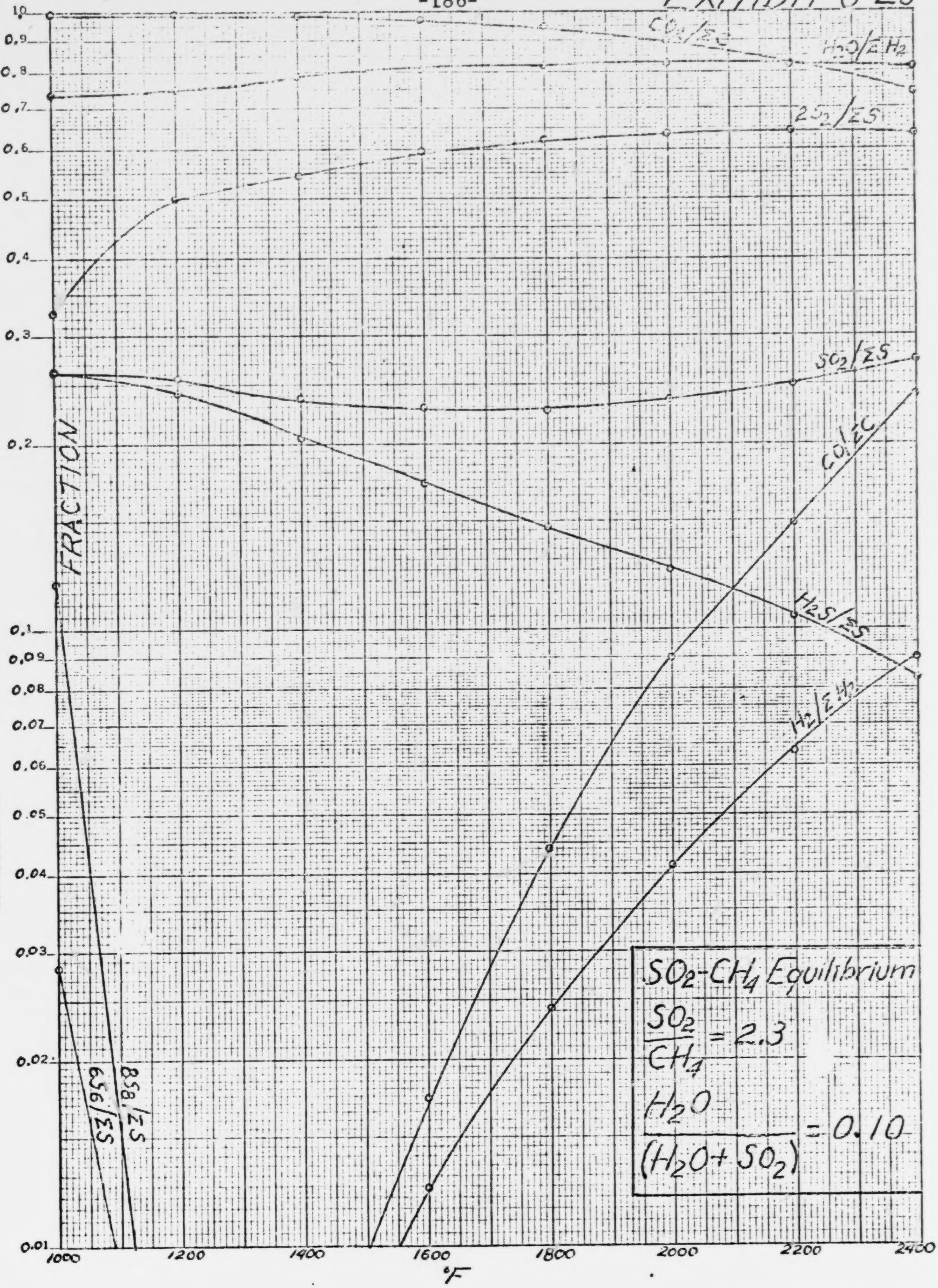
Line No.	Cost Item	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	12.7 MCF	$.30 Avg.	$ 3.81	$ 503.0
1.	Total Raw Materials	-	-	$ 3.81	$ 503.0
2.	Direct Labor	18,000 MH	$3.50	.48	63.0
3.	Supervision	1.	$10,000	.08	10.0
4.	Maintenance	5% FC	-	.93	122.4
5.	Plant Supplies	10%(L2+L3)	-	.05	7.3
	Utilities:				
6. a)	Electricity	40 KWH	$.01	.40	52.8
b)	Cooling Water	4.0 MG	.05	.20	26.4
c)	Boiler Feed Water	MG	.25	.00	6.6
d)	Fuel Gas	1.0 MCF	.30	.30	39.6
7.	Total Utilities	-	-	.90	118.8
8.	Other Direct (Catalyst)	3.8 Lb.	.13	.50	66.0
9.	TOTAL DIRECT	-	-	$ 6.75	$ 890.5
10.	Payroll Burden	30% L2	-	.14	18.9
11.	Plant Overhead	70%(L2+L3)	-	.39	51.1
12.	Pack & Ship	1000 MH	3.90	.03	3.9
13.	Waste Disposal	1000 MH	3.90	.03	3.9
14.	Other Indirect	-	-	-	-
15.	TOTAL INDIRECT	-	-	$ .59	$ 77.8
16.	Depreciation	10% FC	-	1.85	244.7
17.	Taxes	2% FC	-	.37	48.9
18.	Insurance	1% FC	-	.19	24.5
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$ 2.41	$ 318.1
21.	TOTAL OPERATING	-	-	$ 9.75	$1,286.4
22.	Credits LP Steam	1.6 M Lb.	.25	.40	52.8
23.	NET OPERATING	-	-	$ 9.35	$1,233.6

EXHIBIT 8-24



K&E SEMI-LOGARITHMIC 46 5130
2 CYCLES X 140 DIVISIONS MADE IN U.S.A.
KEUFFEL & ESSER CO.

EXHIBIT 8-25



K&E SEMI-LOGARITHMIC 46 5130
2 CYCLES X 140 DIVISIONS MADE IN U.S.A.
KEUFFEL & ESSER CO.

EXHIBIT 8-23

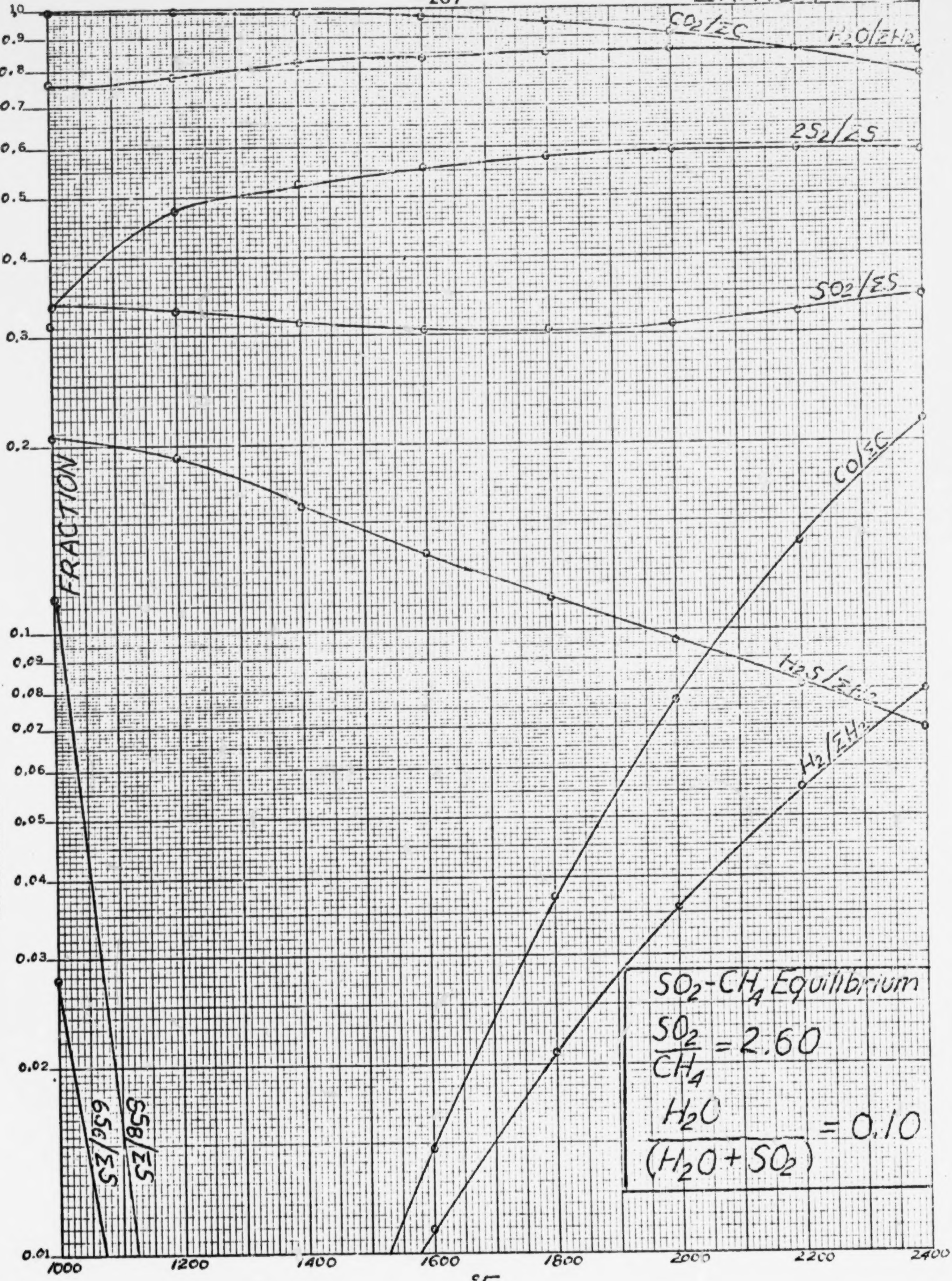
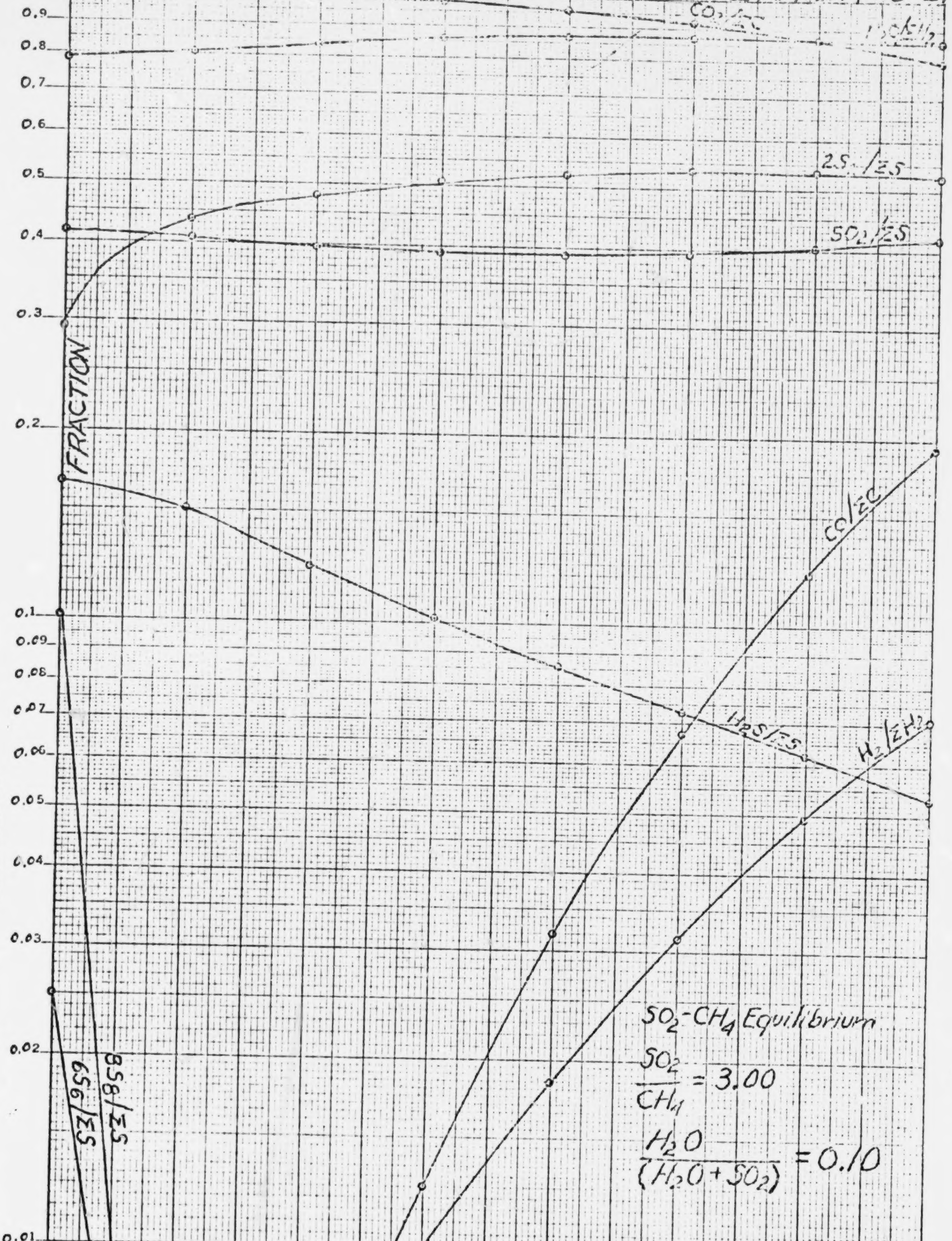

$CO_2/\Sigma C$
$H_2O/\Sigma H_2$
$2S_2/\Sigma S$
$SO_2/\Sigma S$
$CO/\Sigma C$
$H_2S/\Sigma H_2$
$H_2/\Sigma H_2$
$S_6/\Sigma S$
$S_8/\Sigma S$
FRACTION
1.0
0.9
0.8
0.7
0.6
0.5
0.4
0.3
0.2
0.1
0.09
0.08
0.07
0.06
0.05
0.04
0.03
0.02
0.01
1000
1200
1400
1600
1800
2000
2200
2400
°F
SO_2-CH_4 Equilibrium
$\frac{SO_2}{CH_4} = 2.60$
$\frac{H_2O}{(H_2O + SO_2)} = 0.10$
K+E SEMI-LOGARITHMIC 2 CYCLES X 140 DIVISIONS KEUFFEL & ESSER CO.
46 5130 MADE IN U.S.A.

EXHIBIT 8-27



K&E SEMI-LOGARITHMIC 46 5130
2 CYCLES X 140 DIVISIONS MADE IN U.S.A.
KEUFFEL & ESSER CO.

EXH. 8-28

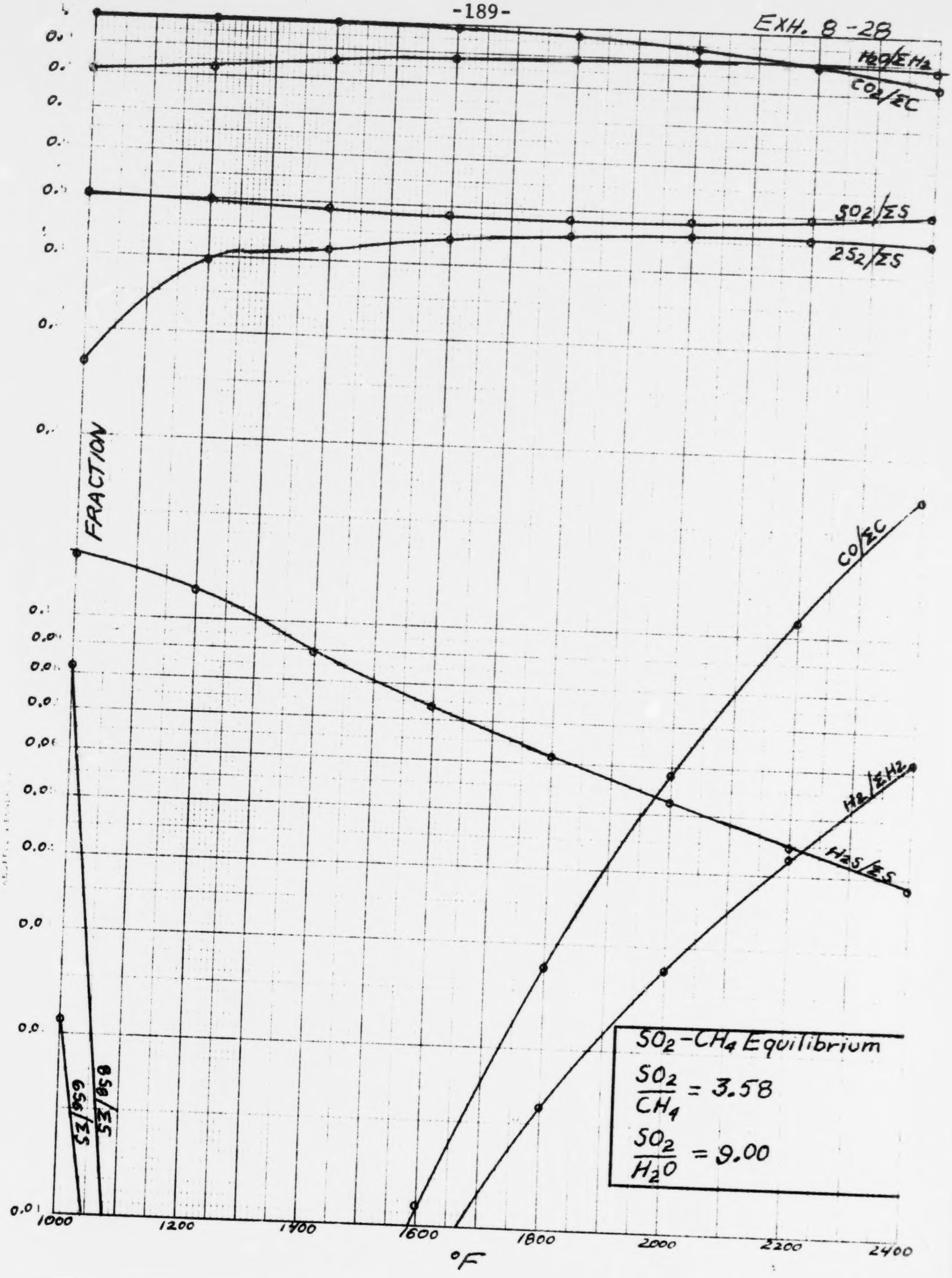

$H_2O/\Sigma H_2$
$CO_2/\Sigma C$
$SO_2/\Sigma S$
$2S_2/\Sigma S$
FRACTION
$CO/\Sigma C$
$H_2/\Sigma H_2$
$H_2S/\Sigma S$
$S_6/\Sigma S$
$S_8/\Sigma S$
0.01
1000
1200
1400
1600
1800
2000
2200
2400
°F
SO_2-CH_4 Equilibrium
$\frac{SO_2}{CH_4} = 3.58$
$\frac{SO_2}{H_2O} = 9.00$

EXHIBIT 8-29

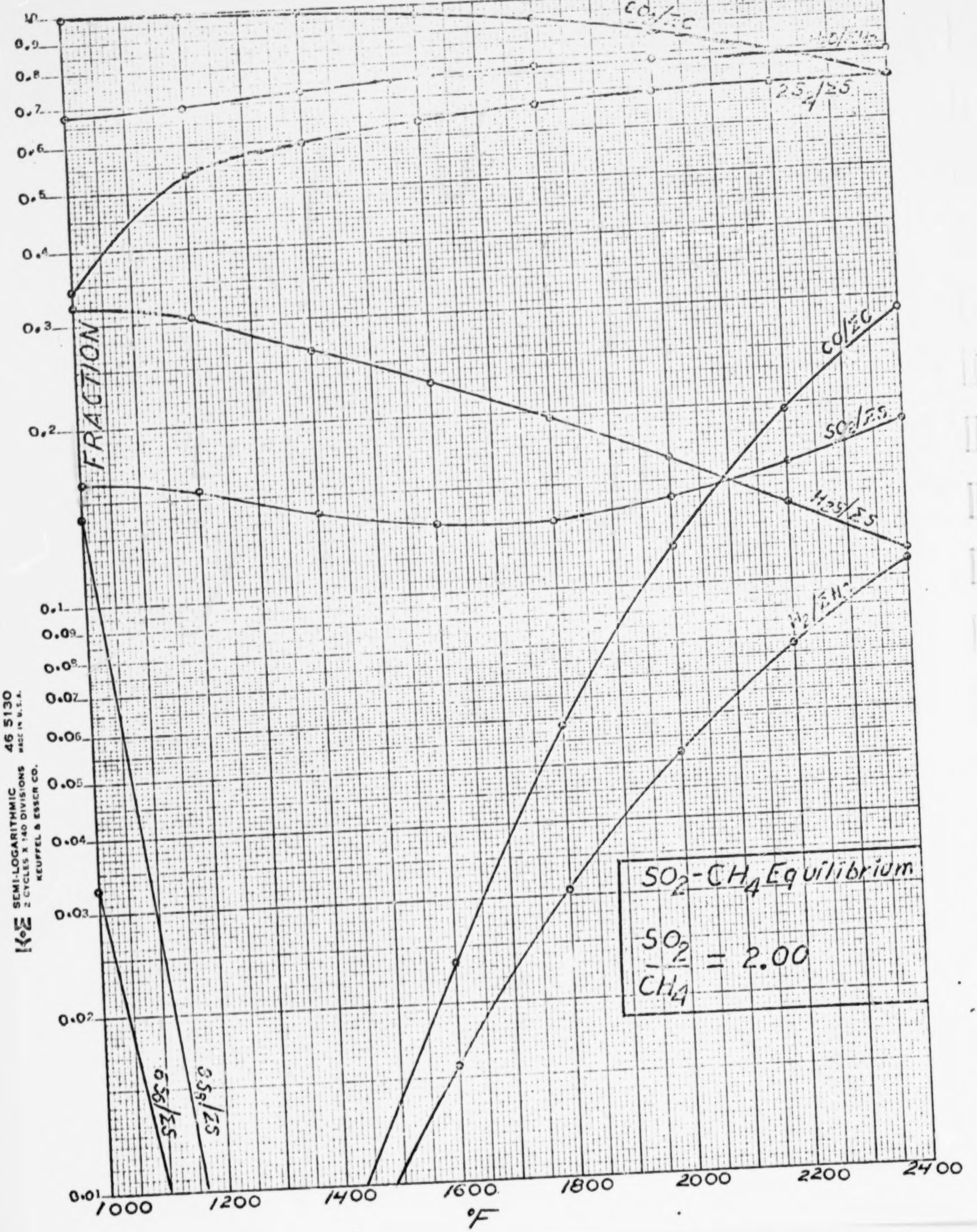
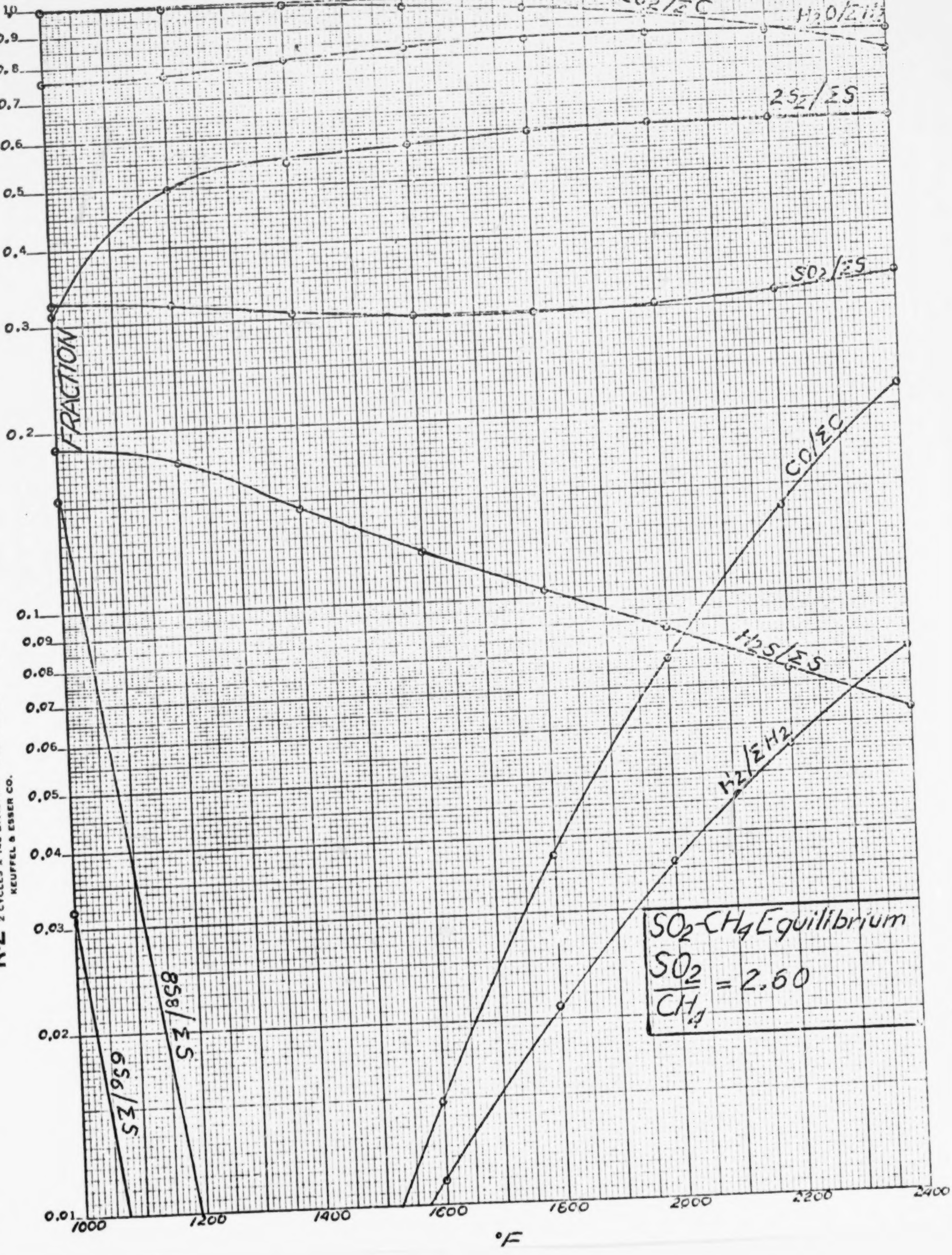
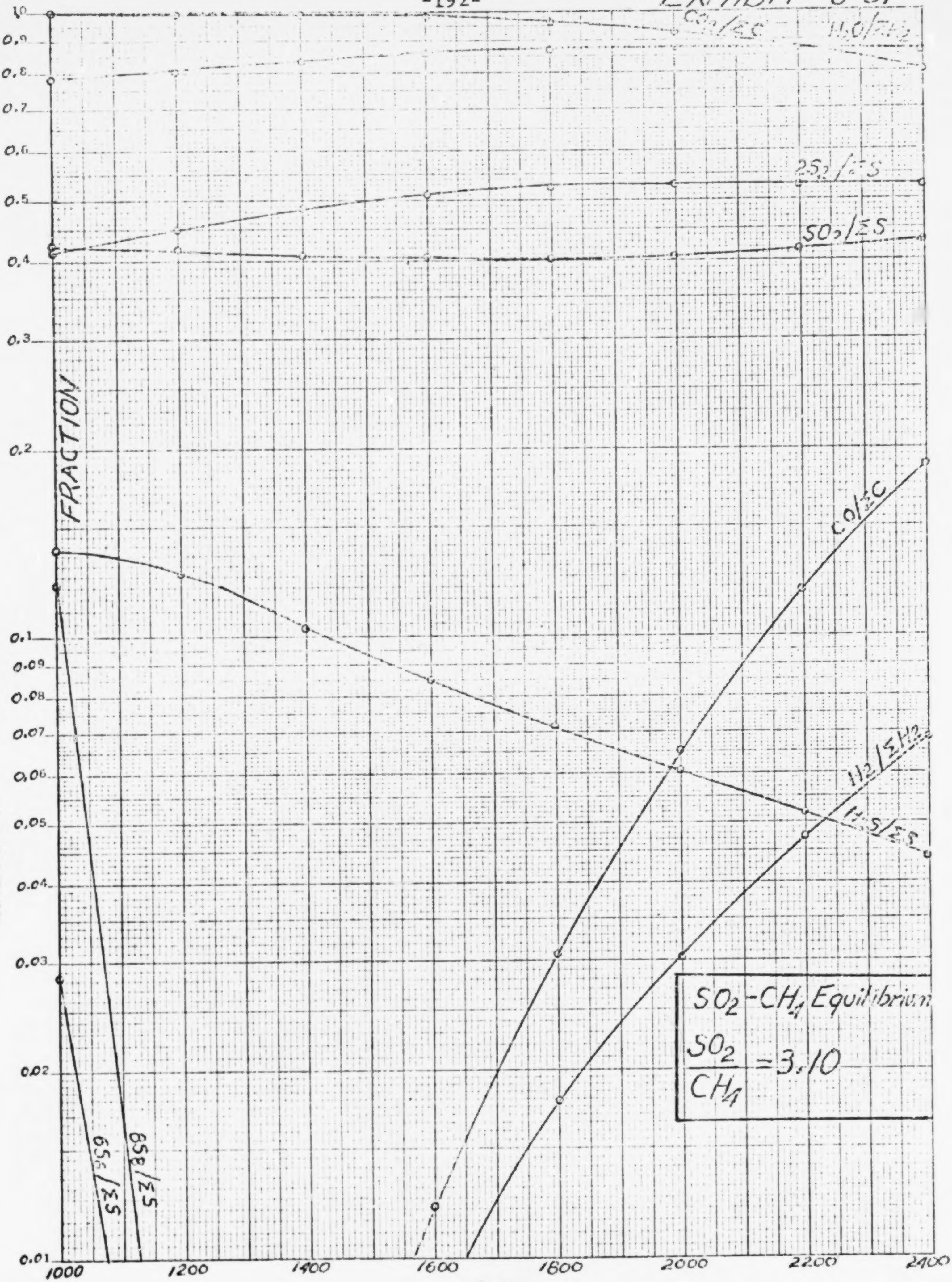

FRACTION
1.0
0.9
0.8
0.7
0.6
0.5
0.4
0.3
0.2
0.1
0.09
0.08
0.07
0.06
0.05
0.04
0.03
0.02
0.01
1000
1200
1400
1600
1800
2000
2200
2400
°F
$CO_2/\Sigma C$
$H_2O/\Sigma H_2$
$2S_2/\Sigma S$
$CO/\Sigma C$
$SO_2/\Sigma S$
$H_2S/\Sigma S$
$H_2/\Sigma H_2$
$8S_8/\Sigma S$
$6S_6/\Sigma S$
SO_2-CH_4 Equilibrium
$\frac{SO_2}{CH_4} = 2.00$
K&E SEMI-LOGARITHMIC 46 5130
2 CYCLES X 140 DIVISIONS MADE IN U.S.A.
KEUFFEL & ESSER CO.

EXHIBIT 8-30



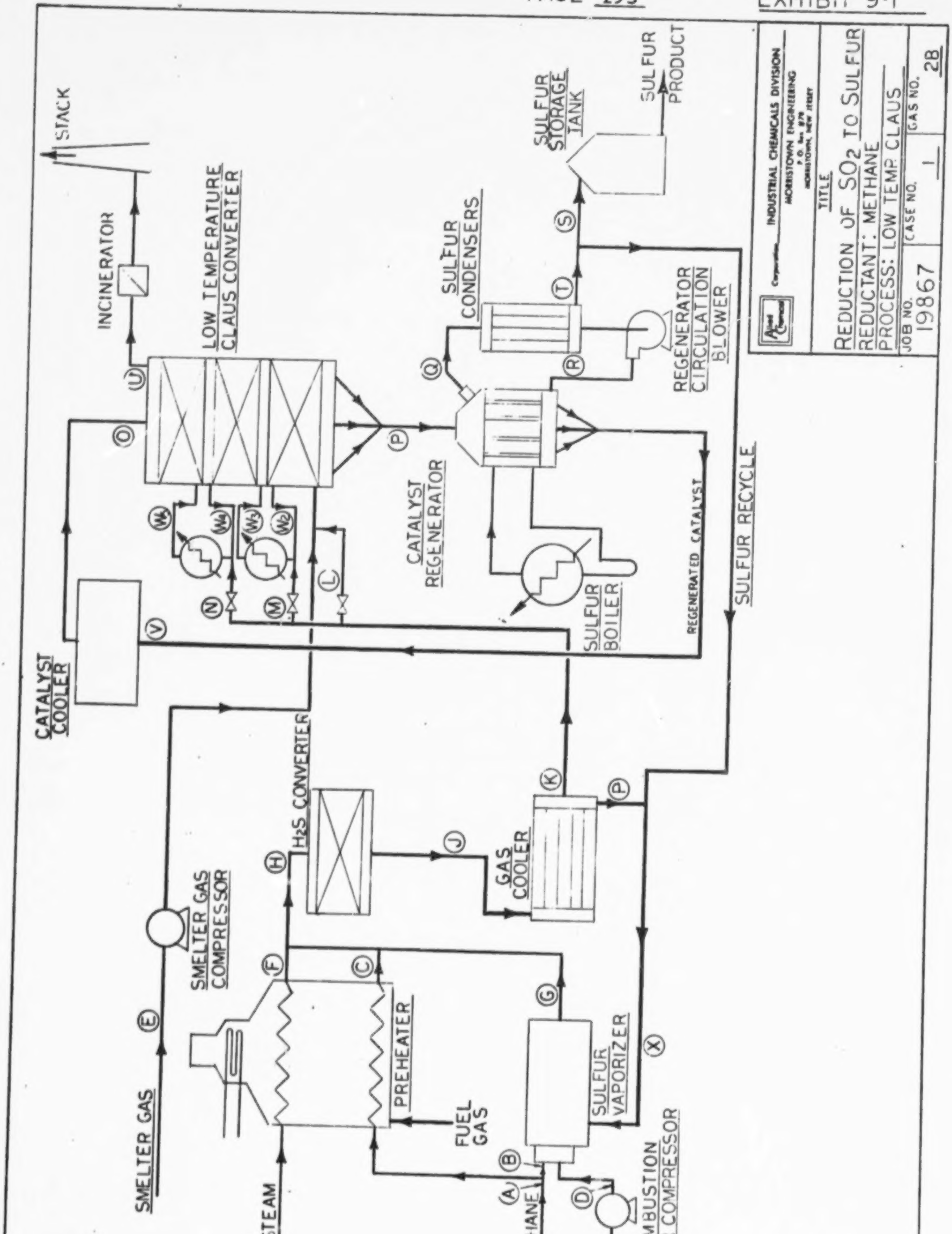
K•E SEMI-LOGARITHMIC 2 CYCLES X 140 DIVISIONS KEUFFEL & ESSER CO. 46 5130 MADE IN U.S.A.

EXHIBIT 8-31



K+E SEMI-LOGARITHMIC 46 5130
2 CYCLES X 140 DIVISIONS MADE IN U.S.A.
KEUFFEL & ESSER CO.

EXHIBIT 9-1


SMELTER GAS
(E)
SMELTER GAS COMPRESSOR
STEAM
PREHEATER
FUEL GAS
(F)
(C)
(H)
H₂S CONVERTER
(J)
GAS COOLER
(K)
(P)
METHANE
(A)
(B)
AIR
(D)
COMBUSTION AIR COMPRESSOR
SULFUR VAPORIZER
(G)
(X)
CATALYST COOLER
(V)
(N)
(M)
(L)
(W₁)
(W₄)
(W₃)
(W₂)
(O)
LOW TEMPERATURE CLAUS CONVERTER
(U)
INCINERATOR
STACK
(P)
CATALYST REGENERATOR
SULFUR BOILER
REGENERATED CATALYST
SULFUR RECYCLE
(Q)
SULFUR CONDENSERS
(R)
REGENERATOR CIRCULATION BLOWER
(T)
(S)
SULFUR STORAGE TANK
SULFUR PRODUCT

Allied Chemical Corporation
INDUSTRIAL CHEMICALS DIVISION
MORRISTOWN ENGINEERING
P. O. Box 879
MORRISTOWN, NEW JERSEY

TITLE
REDUCTION OF SO_2 TO SULFUR
REDUCTANT: METHANE
PROCESS: LOW TEMP. CLAUS

JOB NO.	CASE NO.	GAS NO.
19867	1	2B

DRAWING NO. 37187

LOW TEMPERATURE [illegible]

EXH. 9-2

CASE I 70000 SCFM 23 SMELTER GAS

PROCESS POINTS	A	B	C	D	E	F	G	H	J	K	L	M	N	Q	R	U
TEMP. °C	25	25	705	25	125	625	625	625	720	125	125	125	125	420	125	150
VOL. SCFM	2550	820	1730	7780	70000	1890	11960	15580	15890	15580	6550	6550	2480	1630	514	82440
VOL. CFM@T°	2780	890	6200	8490	102060	6220	39330	51200	7790	22720	9550	9550	3620	4140	750	127700
% CH_4	100.00	100.00	100.00					11.10								
% H_2O						100.00	13.14	22.20	4.48	4.56	4.56	4.56	4.56			8.03
% H_2S									37.58	38.31	38.31	38.31	38.31			0.07
% CO_2							6.58	5.06	11.44	11.65	11.65	11.65	11.65			2.56
% COS									1.94	1.98	1.98	1.98	1.98			0.004
% CO									2.39	2.44	2.44	2.44	2.44			0.46
% S_2							28.90	22.20	1.94							0.00
% H_2									1.49	1.52	1.52	1.52	1.52			0.28
% CS_2									0.08	0.08	0.08	0.08	0.08			0.00
% SO_2					4.50											0.04
% O_2				21.00	16.50											14.01
% N_2				79.00	79.00		51.38	39.44	38.68	39.46	39.46	39.46	39.46	31.50	100.00	74.55
% S_6														37.70		
% S_8														30.80		

POINT	SULFUR DISTRIBUTION		
Ⓣ	FROM CATALYST	600.0	N.T. / DAY
Ⓟ	FROM GAS COOLER	39.5	" "
	TOTAL S RECOVERY	639.5	" "
Ⓧ	RECYCLE SULFUR	443.7	" "
Ⓢ	NET RECOVERY	195.8	" "

CONVERSION

EQUILIBRIUM 99.5 %
USED HERE 99.0 %

SULFUR YIELD 96.87 %

CATALYST CIRCULATION RATE 100.0 NT/Hr.

METHANE FACTOR 18750 SCF/TON SULFUR

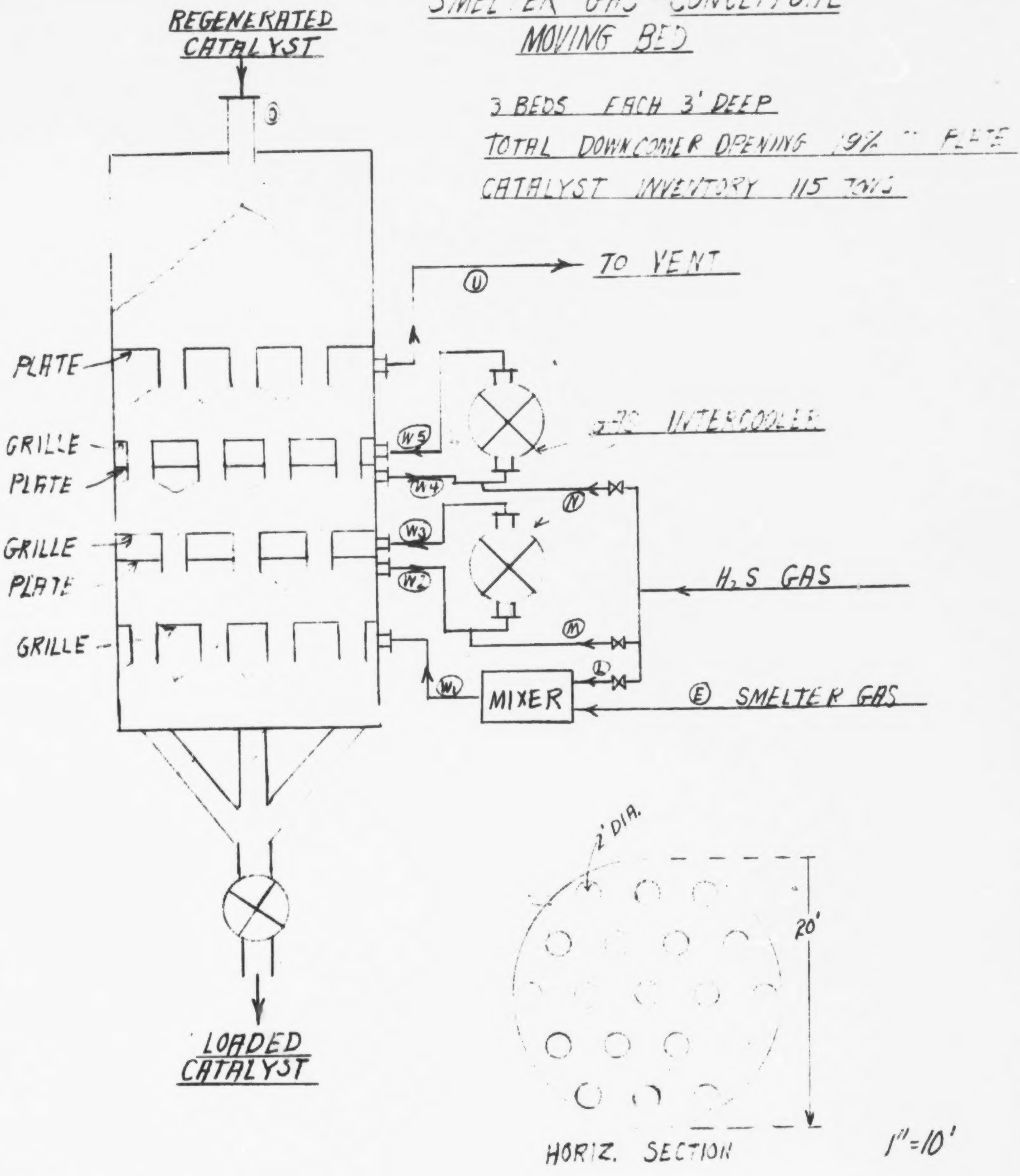
PROCESS POINTS	TEMP. °C
Q	125
V	400
W_1	125
W_2	189
W_3	125
W_4	189
W_5	125

EXHIBIT 9-2

EXH 9-3

CASE 1

LOW TEMPERATURE CLAUS CONVERTER SMELTER GAS - CONCEPTUAL MOVING BED



3 BEDS EACH 3' DEEP

TOTAL DOWNCOMER OPENING 19% OF PLATE

CATALYST INVENTORY 115 TONS

REGENERATED CATALYST
Ⓢ
TO VENT
Ⓤ
PLATE
GRILLE
PLATE
GRILLE
PLATE
GRILLE
W5
W4
W3
W2
W1
INTERCOOLER
Ⓝ
Ⓜ
Ⓛ
MIXER
H_2S GAS
Ⓔ SMELTER GAS
LOADED CATALYST
2' DIA.
20'
HORIZ. SECTION
1" = 10'

CATALYST REGENERATOR

EXH 9-4

CASE I

INDIRECT HEAT
MOVING BED

CATALYST INVENTORY 50.0 TONS
0.5 Hr RESIDENCE

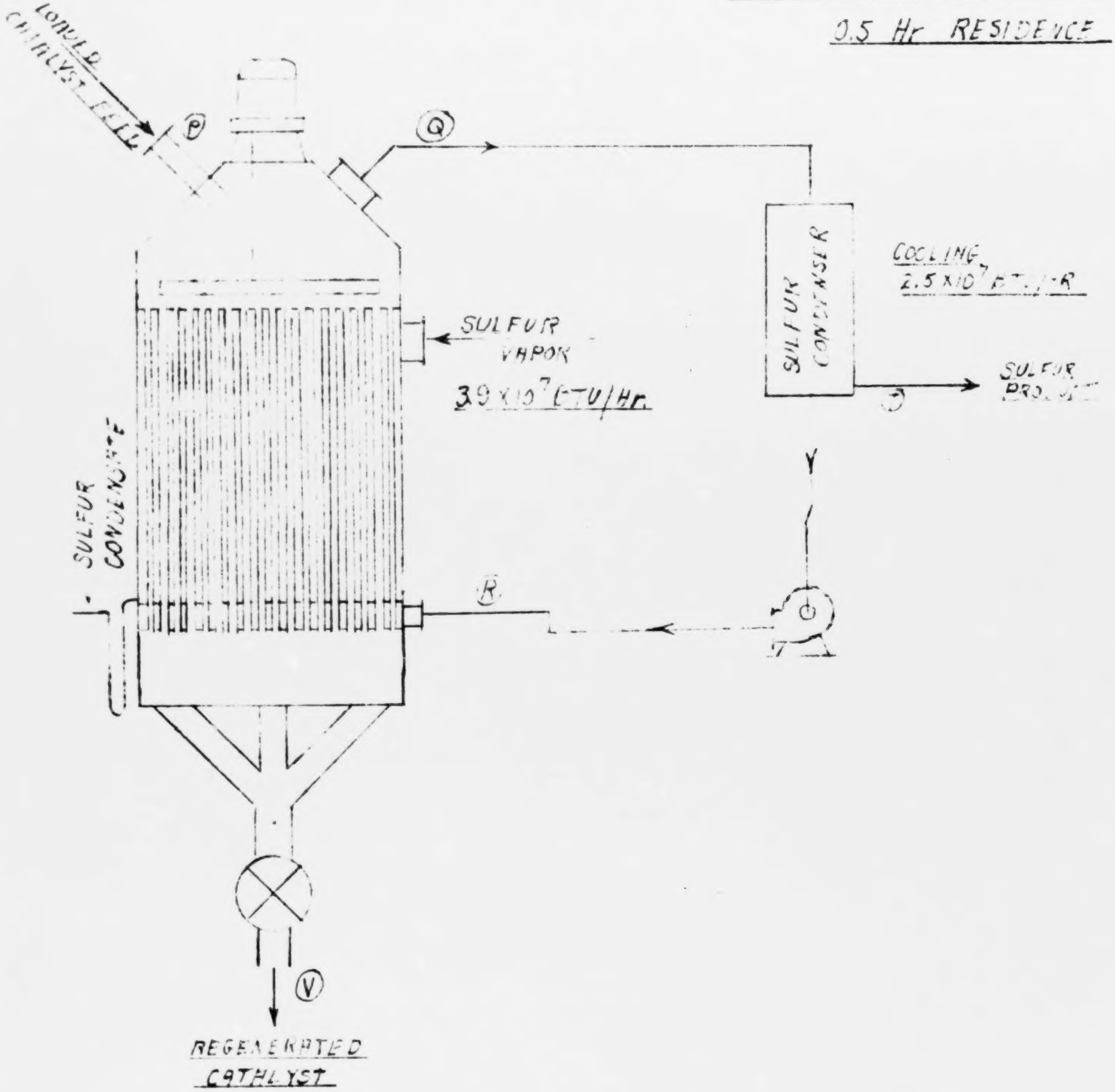
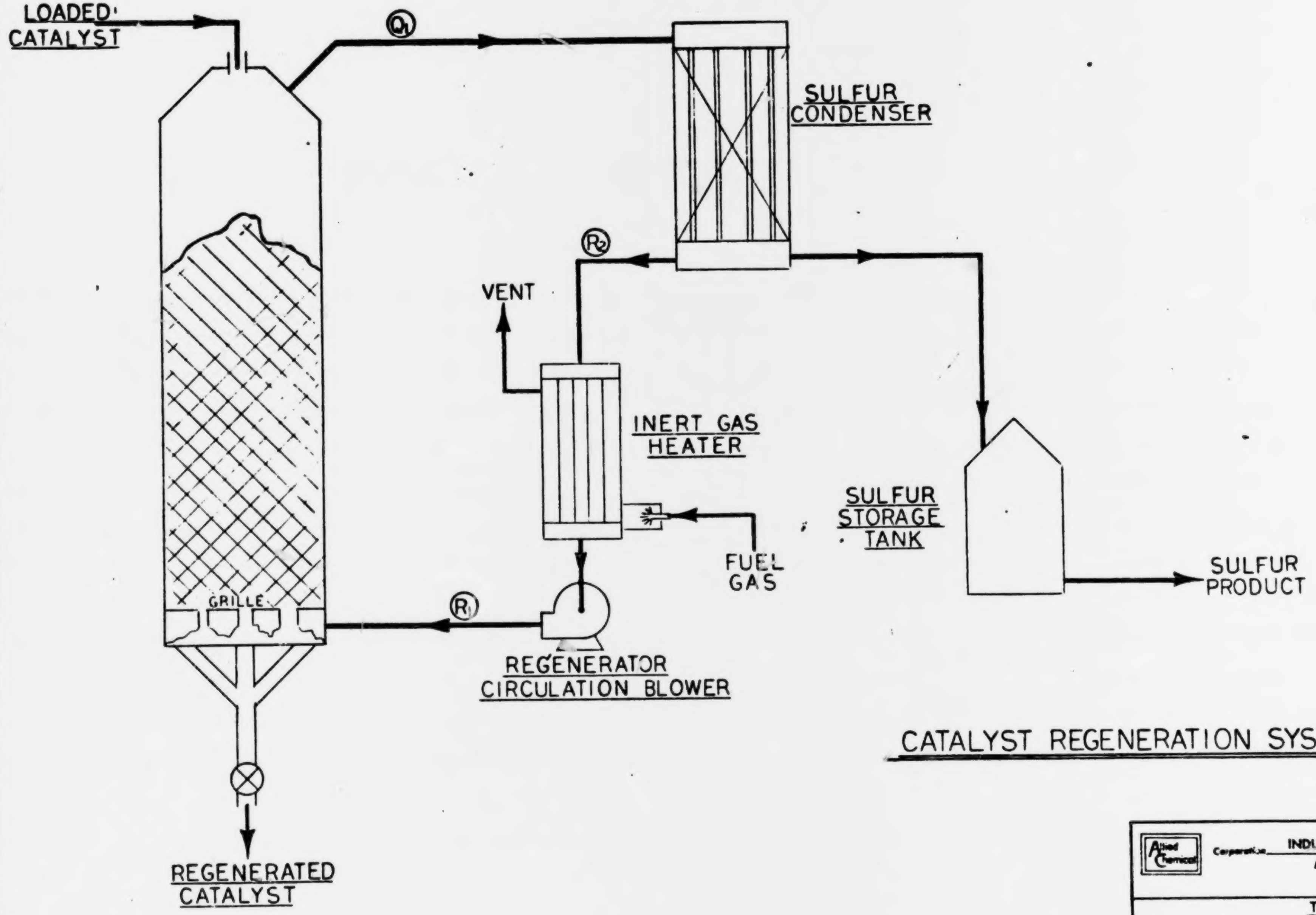

LOADED CATALYST FEED
P
Q
SULFUR CONDENSER
COOLING 2.5×10^7 BTU/HR
SULFUR PRODUCT
S
Y
SULFUR VAPOR
3.9×10^7 BTU/Hr.
SULFUR CONDENSATE
R
V
REGENERATED CATALYST

EXHIBIT 9-5


LOADED CATALYST
Q1
SULFUR CONDENSER
R2
VENT
INERT GAS HEATER
SULFUR STORAGE TANK
FUEL GAS
SULFUR PRODUCT
GRILLE
R1
REGENERATOR CIRCULATION BLOWER
REGENERATED CATALYST

CATALYST REGENERATION SYSTEM ONLY

Allied Chemical Corporation INDUSTRIAL CHEMICALS DIVISION
MORRISTOWN ENGINEERING
P.O. Box 279
MORRISTOWN, NEW JERSEY

TITLE
REDUCTION OF SO_2 TO SULFUR
REDUCTANT: METHANE
PROCESS: LOW TEMP CLAUS

JOB NO.	CASE NO.	GAS NO.
19867	1A	2B

DRAWING NO. 37188

EXH. 9-6

H_2S SYNTHESIS

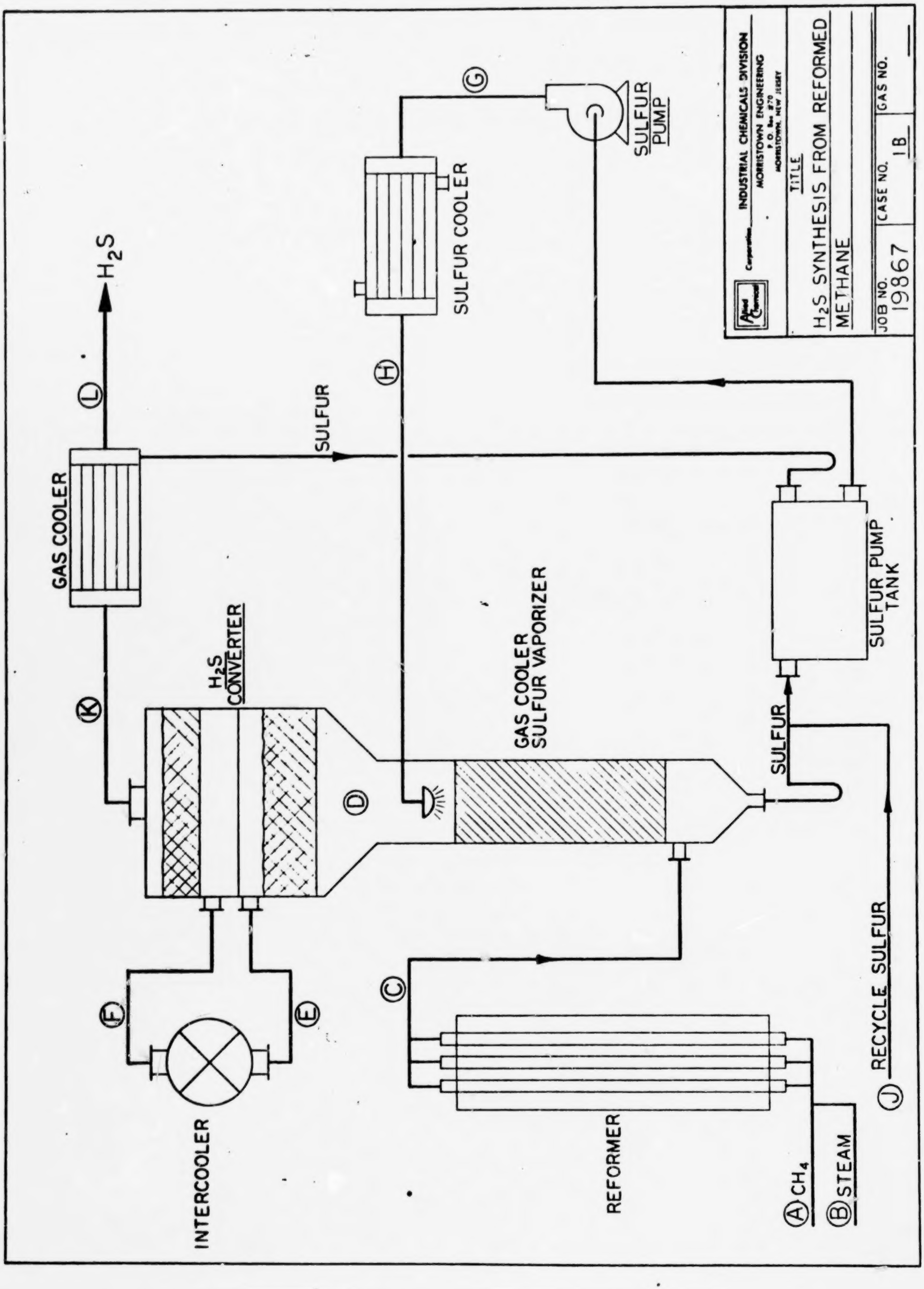
CASE 1-B FROM REFORMED METHANE

EQUIVALENT TO 70000 SCFM 2B SMELTER GAS

PROCESS POINTS	A	B	C	D	E	F	G	H	K	L
TEMP. °C	25	205	898	300	625	300	300	250	540	125
VOL. SCFM	1590	4770	9470	10270	10700	10700			9640	9610
VOL. CFM @ T°C	1730	8350	40620	21560	35190	22460			28710	14010
% CH_4	100.00		0.35	0.32	0.11	0.11				
% H_2O		100.00	29.03	26.78	19.03	19.03			16.74	16.80
% CO			11.57	10.68	3.96	3.96			0.11	0.11
% H_2			54.20	50.00	19.05	19.05			1.31	1.31
% CO_2			4.85	4.48	10.53	10.53			16.17	16.26
% S_8				7.74						
% H_2S					35.69	35.69			64.98	65.18
% COS					0.15	0.15			0.34	0.34
% S_2					11.48	11.48			0.34	0.00

RECYCLE SULFUR 408.1 TONS/DAY

EXHIBIT 9-7


INTERCOOLER
F
E
REFORMER
Ⓐ CH$_4$
Ⓑ STEAM
C
Ⓙ RECYCLE SULFUR
H_2S CONVERTER
D
K
GAS COOLER
L
H_2S
SULFUR
GAS COOLER SULFUR VAPORIZER
SULFUR
SULFUR PUMP TANK
H
SULFUR COOLER
G
SULFUR PUMP

Allied Chemical Corporation
INDUSTRIAL CHEMICALS DIVISION
MORRISTOWN ENGINEERING
P. O. Box #70
MORRISTOWN, NEW JERSEY

TITLE
H_2S SYNTHESIS FROM REFORMED METHANE

JOB NO.	CASE NO.	GAS NO.
19867	1B	—

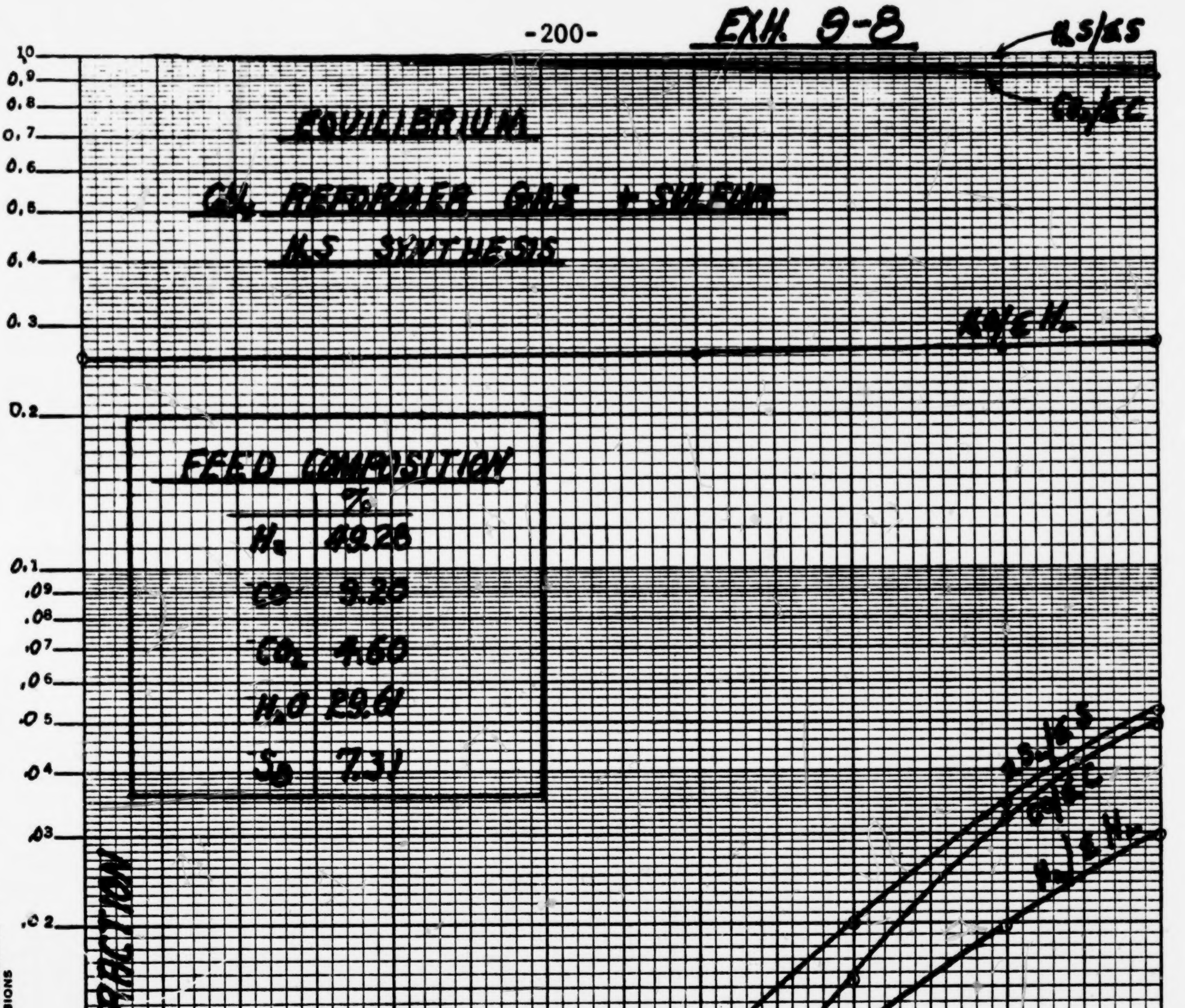
DRAWING NO. 37189

EXH. 9-8

EQUILIBRIUM

CH_4 REFORMER GAS + SULFUR

H_2S SYNTHESIS

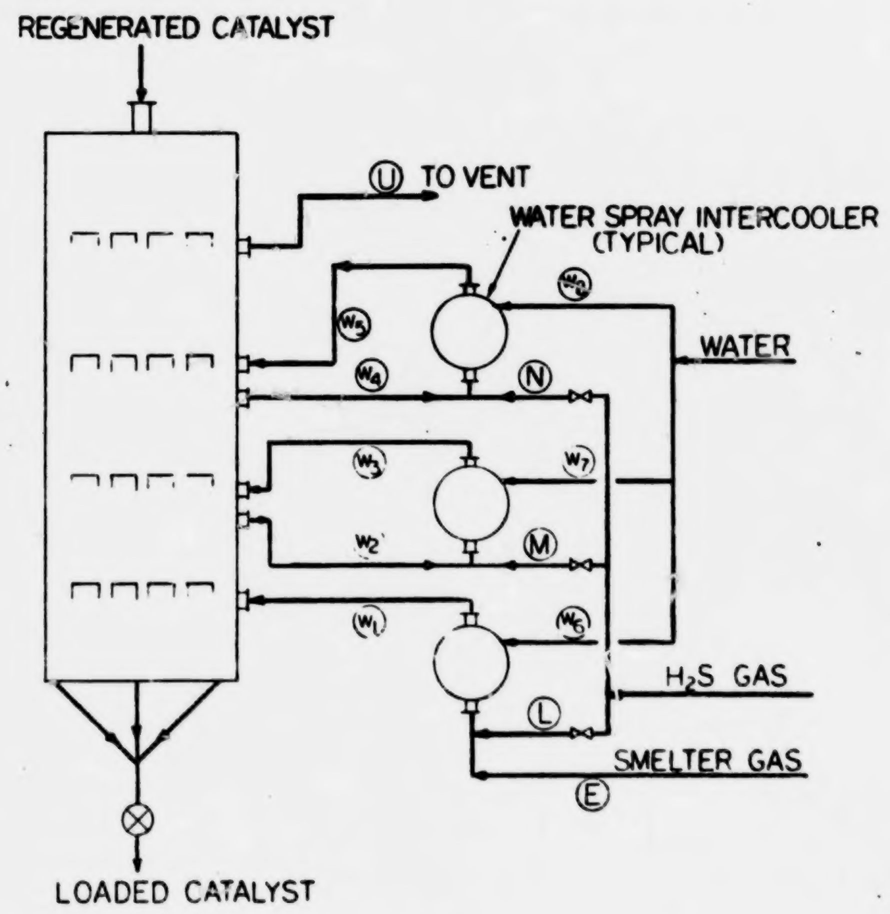


FEED COMPOSITION

	%
H_2	49.28
CO	9.20
CO_2	4.60
H_2O	29.61
S_8	7.31

K+E SEMI-LOGARITHMIC 358-71
KEUFFEL & ESSER CO. MADE IN U.S.A.
3 CYCLES X 70 DIVISIONS

EXHIBIT 9-9


REGENERATED CATALYST
U
TO VENT
WATER SPRAY INTERCOOLER (TYPICAL)
W_5
W_4
N
WATER
W_3
W_7
W_2
M
W_1
W_6
H_2S GAS
L
SMELTER GAS
E
LOADED CATALYST

PROCESS POINTS	E	L	M	N	W_1	W_2	W_3	W_4	W_5	U
TEMP °C	125	125	125	125	65.0	116.0	65.0	116.0	65.0	116.0
VOL SCFM	70,000	5190	5190	5190	77410	76390	84900	83780	92400	91380
DEW POINT °C					31.8		42.0		48.9	

EQUILIBRIUM CONVERSION · 99.76 %

INJECTION COOLING WATER
GAL./MIN.
W_6 - 22.3
W_7 - 20.8
W - 23.0

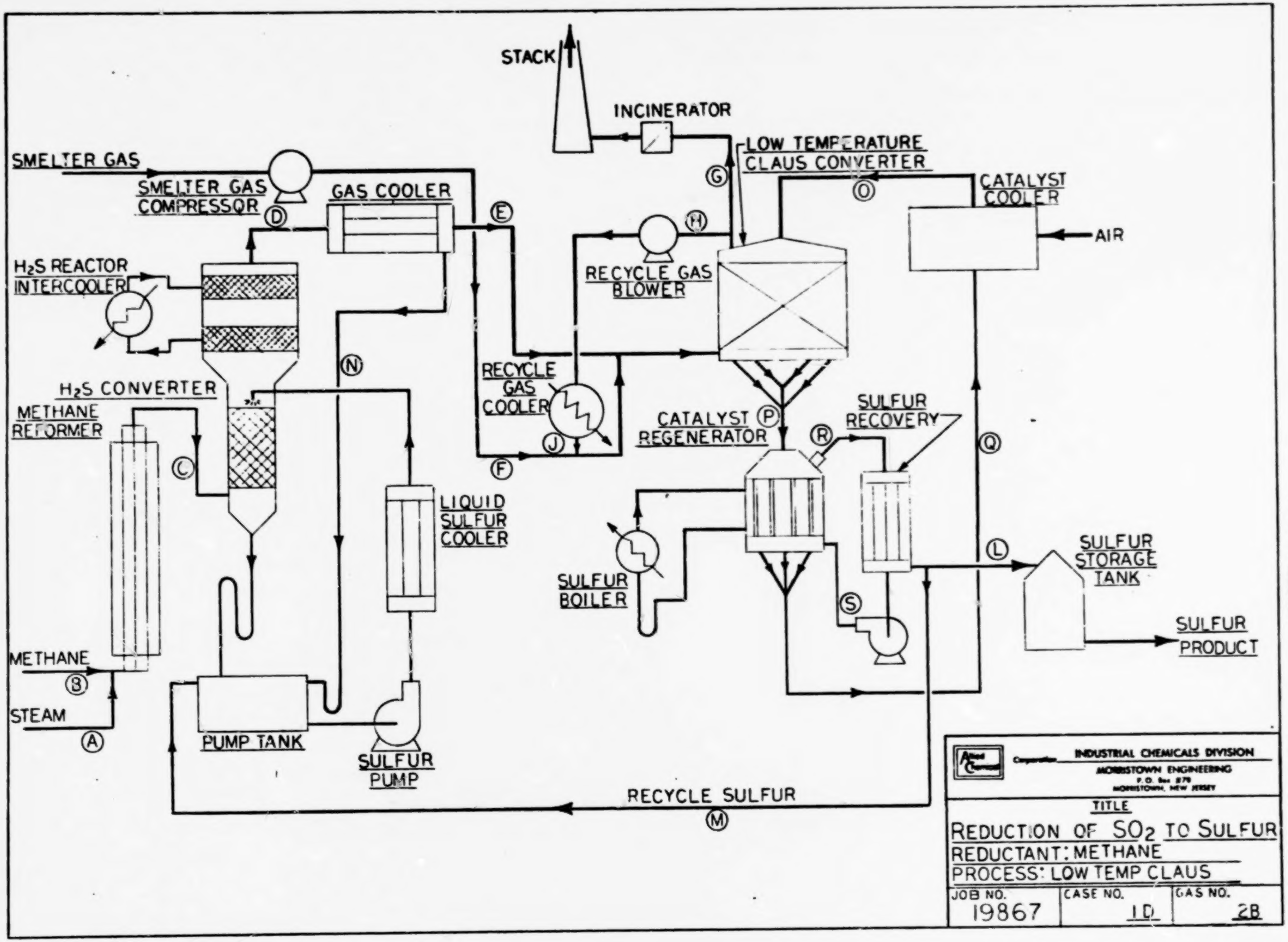
Allied Chemical Corporation INDUSTRIAL CHEMICALS DIVISION
MORRISTOWN ENGINEERING
P. O. Box #70
MORRISTOWN, NEW JERSEY

TITLE
REDUCTION OF SO_2 TO SULFUR
REDUCTANT: H_2S
PROCESS: LOW TEMP. CLAUS

JOB NO.	CASE NO.	GAS NO.
19867	PART OF 1 C	2-B

DRAWING NO. 37190

EXHIBIT 9-10


STACK
INCINERATOR
SMELTER GAS
SMELTER GAS COMPRESSOR
D
GAS COOLER
E
G
LOW TEMPERATURE CLAUS CONVERTER
O
CATALYST COOLER
AIR
H
RECYCLE GAS BLOWER
H₂S REACTOR INTERCOOLER
N
H₂S CONVERTER
METHANE REFORMER
C
RECYCLE GAS COOLER
J
F
CATALYST REGENERATOR
P
SULFUR RECOVERY
R
Q
LIQUID SULFUR COOLER
SULFUR BOILER
L
SULFUR STORAGE TANK
S
SULFUR PRODUCT
METHANE
B
STEAM
A
PUMP TANK
SULFUR PUMP
RECYCLE SULFUR
M

Allied Chemical Corporation INDUSTRIAL CHEMICALS DIVISION
MORRISTOWN ENGINEERING
P.O. Box 279
MORRISTOWN, NEW JERSEY

TITLE

REDUCTION OF SO_2 TO SULFUR
REDUCTANT: METHANE
PROCESS: LOW TEMP CLAUS

JOB NO.	CASE NO.	GAS NO.
19867	1D	2B

DRAWING NO. 37191

EXHIBIT 9-11

LOW TEMPERATURE CLAUS

EXH. 9-11
CASE 1-D

ADIABATIC REACTOR WITH VENT GAS RECYCLE
70000 SCFM 2B SMELTER GAS

PROCESS POINTS	A	B	C	D	E	F	G	H	J	O	P	Q
TEMP. °C	205	25	898	540	125	100	176	176	100	30	100	400
VOL. SCFM	4770	1590	9470	9640	9610	70000	76510	70000	70000			
VOL. CFM @ T°	8350	1730	40620	28710	14010	95640	125830	115120	95640			
% CH_4		100.0	0.35							COOLED REGEN. CATALYST	LOADED CATALYST	REGEN. CATALYST
% H_2O	100.00		29.03	16.74	16.80		5.31	5.31	5.31			
% H_2S				64.98	65.18		0.06	0.06	0.06			
% CO_2			4.85	16.17	16.26		1.09	1.09	1.09			
% COS				0.34	0.34		0.00	0.00	0.00			
% CO			11.57	0.11	0.11		0.01	0.01	0.01			
% H_2			54.20	1.31	1.31		0.09	0.09	0.09			
% SO_2						4.50	0.03	0.03	0.03			
% S_2				0.34	0.00							
% O_2						16.50	17.92	17.92	17.92			
% N_2						79.00	75.49	75.49	75.49			

CONVERSION

EQUILIBRIUM 99.0 %
USED HERE 98.5 %

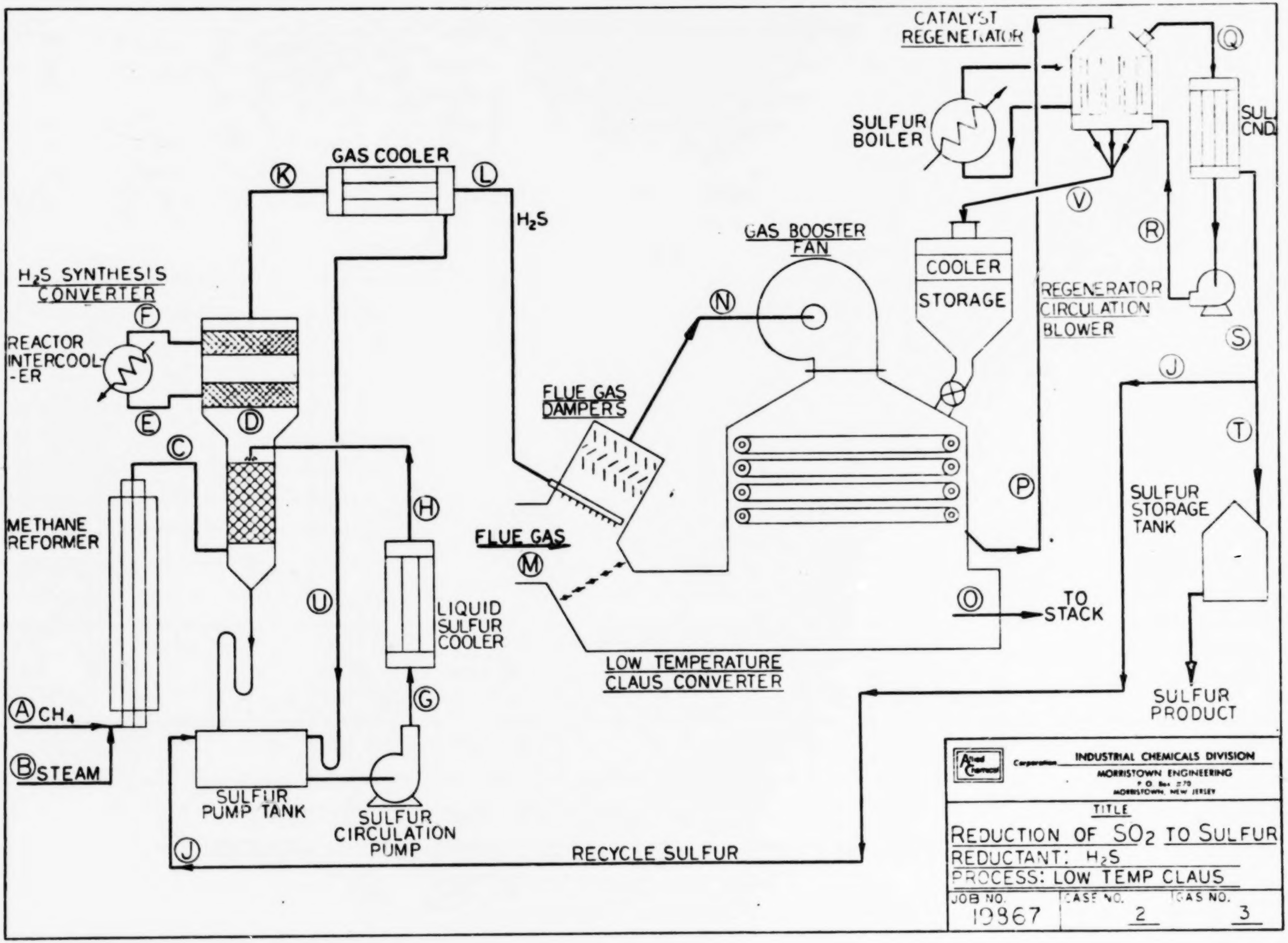
POINT	SULFUR DISTRIBUTION	
Ⓟ	FROM CATALYST	601.3 TON/DAY
Ⓝ	FROM GAS COOLER	4.2 "
	TOTAL S RECOVERY	605.5 "
Ⓜ	RECYCLE SULFUR	408.1 "
Ⓛ	NET S RECOVERY	197.4 "

SULFUR YIELD 97.7 %

CATALYST CIRCULATION RATE 100 TONS/Hr.

METHANE FACTOR 11600 SCF/TON SULFUR
H_2S FROM REFORMED METHANE

EXHIBIT 9-12


GAS COOLER
K
L
H_2S
H_2S SYNTHESIS CONVERTER
F
REACTOR INTERCOOL-ER
E
D
C
METHANE REFORMER
A CH_4
B STEAM
U
H
LIQUID SULFUR COOLER
G
SULFUR PUMP TANK
SULFUR CIRCULATION PUMP
J
RECYCLE SULFUR
FLUE GAS DAMPERS
FLUE GAS
M
N
GAS BOOSTER FAN
LOW TEMPERATURE CLAUS CONVERTER
O
TO STACK
P
COOLER
STORAGE
CATALYST REGENERATOR
SULFUR BOILER
V
R
REGENERATOR CIRCULATION BLOWER
S
Q
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J
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SULFUR STORAGE TANK
SULFUR PRODUCT

Allied Chemical Corporation INDUSTRIAL CHEMICALS DIVISION
MORRISTOWN ENGINEERING
P. O. Box [illegible]
MORRISTOWN, NEW JERSEY

TITLE
REDUCTION OF SO_2 TO SULFUR
REDUCTANT: H_2S
PROCESS: LOW TEMP CLAUS

JOB NO.	CASE NO.	GAS NO.
19867	2	3

DRAWING NO. 37192

LOW TEMPERATURE CLAUS

2.35 MM SCFM #3 STACK GAS

EXH. 9-13

CASE 2

PROCESS POINTS	A	B	C	D	E	F	G	H	K	L	M	N	O	Q	R	P	V
TEMP. °C	25	205	898	300	625	300	300	250	540	125	125	125	138	420	125	138	400
VOL. SCFM	3560	10670	21280	23000	23970	23970			21580	21500	2.35 MM	2.371 MM	2.365 MM	3500	1100		
VOL CFM @ T °C	3880	18700	91000	48290	78830	50310			64310	31380	3.42 MM	3.46 MM	3.56 MM	8900	2370		
% CH_4	100.0		0.35	0.32	0.11	0.11											
% H_2O		100.0	29.03	26.78	19.03	19.03			16.74	16.80	7.25	7.33	7.92				REGENERATED CATALYST
% H_2S					35.69	35.69			64.98	65.18		0.60	0.03			LOADED CATALYST	
% CO_2			4.85	4.48	10.53	10.53			16.17	16.26	14.70	14.71	14.76				
% COS					0.15	0.15			0.34	0.34							
% CO			11.57	10.68	3.96	3.96			0.11	0.11							
% S_2					11.48	11.48			0.34	0.00							
% H_2			54.20	50.00	19.05	19.05			1.31	1.31		0.01	0.01				
% CS_2																	
% SO_2											0.30	0.30	0.015				
% O_2											2.80	2.77	2.78				
% N_2											74.95	74.28	74.48	31.50	100.0		
% S_6														37.70			
% S_8				7.74	0.00	0.00			0.00	0.00				30.80			

POINT	SULFUR DISTRIBUTION		
Ⓢ	FROM CATALYST	1289.4	M.T. /DAY
Ⓤ	FROM GAS COOLER	9.5	" "
	TOTAL S RECOVERY	1298.9	" "
Ⓙ	RECYCLE SULFUR	914.2	" "
Ⓣ	NET S. RECOVERY	384.7	" "

SO_2 CONVERSION

EQUILIBRIUM 96.43 %

USED HERE 95.00 %

SULFUR YIELD 85.02 %

CATALYST CIRCULATION RATE 214.9 M.T. /Hr.

METHANE FACTOR 13300 SCF /TON SULFUR

EXHIBIT 9-13

Figure 9-1

NOT REPRODUCIBLE

LOW TEMPERATURE CLAUS CONVERTER FOR STACK GAS

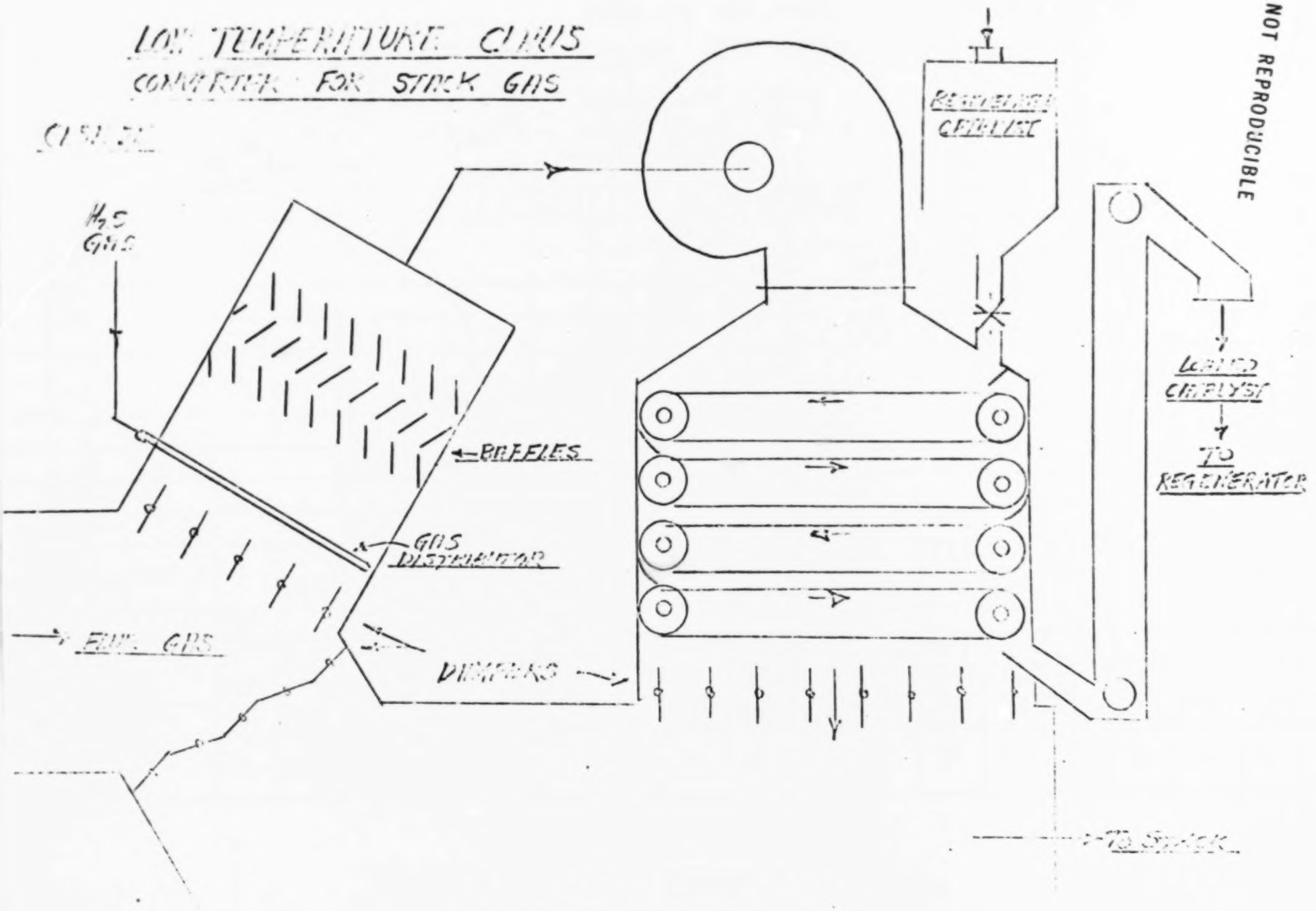

H_2S GAS
BAFFLES
GAS DISTRIBUTOR
FLUE GAS
DAMPERS
REGENERATED CATALYST
LOADED CATALYST
TO REGENERATOR
TO STACK

EXHIBIT 9-15
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: H_2S (Low Temp. Claus) CASE NO. 1
PROCESS: Indirect Regen. Heat GAS NO. 2B

Fixed Capital, Battery Limits Basis $6,070,000
Sulfur Production 196 NTD x 330 Days = 65,000 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	12.7 MCF	$.30 Avg.	$ 3.81	$ 248.0
1.	Total Raw Materials	-	-	$ 3.81	$ 248.0
2.	Direct Labor	18,000 MH	$3.50	.97	63.0
3.	Supervision	1.	$10,000	.15	10.0
4.	Maintenance	5% FC	-	4.67	303.5
5.	Plant Supplies	10%(L2+L3)	-	.11	7.3
	Utilities:				
6. a)	Electricity	260 KWH	$.01	2.60	169.0
b)	Cooling Water	4.4 MG	.05	.22	14.3
c)	Boiler Feed Water	.72 MG	.25	.18	11.7
d)	Fuel Gas	23.3 MCF	.30	7.00	455.0
7.	Total Utilities	-	-	10.00	650.0
8.	Other Direct (Catalyst)	77.0 Lbs.	.13	10.00	650.0
9.	TOTAL DIRECT	-	-	$ 29.71	$ 1,931.8
10.	Payroll Burden	30% L2	-	.29	18.9
11.	Plant Overhead	70%(L2+L3)	-	.79	51.1
12.	Pack & Ship	1000 MH	3.90	.06	3.9
13.	Waste Disposal	1000 MH	3.90	.06	3.9
14.	Other Indirect	-	-	-	-
15.	TOTAL INDIRECT	-	-	$ 1.20	$ 77.8
16.	Depreciation	10% FC	-	9.35	607.0
17.	Taxes	2% FC	-	1.87	121.4
18.	Insurance	1% FC	-	.94	60.7
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$ 12.16	$ 789.1
21.	TOTAL OPERATING	-	-	$ 43.07	$ 2,798.7
22.	Credits 250 PSI Steam	1.7 M Lbs.	.50	.85	55.2
23.	NET OPERATING	-	-	$ 42.22	$ 2,743.5

EXHIBIT 9-16

LOW TEMPERATURE CLAUS
H_2S REDUCTION OF SO_2
2B SMELTER GAS

CASE 1

SUMMATION OF MAJOR EQUIPMENT

1.	Combustion Air Compressor - single stage centrifugal compressor to handle 8000 SCFM at an inlet temperature of 77°F. with a ΔP of 6 PSI. Equipped with a 300 HP motor.	$ 45,000
2.	Sulfur Vaporizer - submerged combustion heater 50 MM BTU/HR. to vaporize 460 NT sulfur/day to exit at 1300°F, burning fuel gas or methane at controlled excess air conditions. A small amount of SO_2 will be generated by this unit for later recovery.	$ 50,000
3.	Smelter Gas Compressor - single stage centrifugal compressor to handle 70,000 SCFM at an inlet temperature of 77°F. with a ΔP of 3 PSI. Equipped with a 1500 HP motor.	$100,000
4.	Preheater - vertical bottom fired process heater, radiant heat duty 7.1 MM BTU/HR., convection heat duty 2 MM BTU/HR., including economizer. Net heat release 14.5MM BTU/HR.	$ 60,000
5.	Hydrogen Sulfide Synthesis Convertors - two 13 foot diameter shop fabricated vertical vessels packed to a depth of 10 feet with alumina type catalyst.	$100,000
6.	Sulfur Storage Tank - vertical steel tank 2,000 ton capacity with steam coils and external insulation.	$ 40,000
7.	Gas Cooler - two stage gas cooling heat exchanger. First stage generating 250 PSIG steam reducing gas temperature to 530°F. Second stage using cooling water to reduce gas temperature to 257°F. Approximate steam generation 14,000 lb./hr.	$ 70,000
8.	Low Temperature Claus Converter - moving catalyst bed design per exhibit 3. Chamber is steel, 20 ft. diameter x 50 ft. high straight side having three catalyst zones with internal downcomers and gas distribution system. Approximately 90% of gas flow is assumed to bypass through each interzone heat exchanger to control bed temperature rise. H_2S is added at several points to prevent excessive temperature rise in bed. Cooling medium for interzone gas coolers is tempered cooling water. Tube wall temperature is held at 250°F. minimum to minimize sulfur deposition on heat exchange surfaces. Including intercoolers, catalyst and auxiliaries.	$750,000

9.	Catalyst Regenerator - indirect heated moving bed design comprising 22 ft. diameter x 50 ft. high carbon steel shell with 380 - 6" tubes 27' long through which catalyst passes countercurrent to an inert or reducing circulating gas stream. Catalyst is distributed above the upper tube sheet with a rake or similar mechanical distributor. Heat for sulfur stripping is provided by condensing sulfur vapor on external tube surfaces. Including catalyst and auxiliaries.----------	$320,000
10.	Catalyst Cooler - Carbon steel 15 ft. diameter by 20 ft. high shell with air distribution/catalyst support grates. Tempered air cools 100 NT per hour catalyst from 790°F. to 260°F. Complete with catalyst, catalyst handling equipment, air tempering coils, air blower with 100 HP motor and auxiliaries.	$ 80,000
11.	Regenerator Gas Exchanger - shell and tube carbon steel heat exchanger designed to reheat circulating stripping gas stream to 700°F. by exchange against regenerator overhead stream----	$ 5,000
12.	Sulfur Coolers and Condensers - 35 PSIG steam generators designed to condense and recover 600 NT/day sulfur from regenerator gas stripping system. Carbon steel construction.	$100,000
13.	Regenerator Circulation Blower - centrifugal single stage blower of forged steel casing construction suitable for service on H_2S or nitrogen with ΔP of 8 PSI complete with 400 HP motor.	$ 90,000
14.	Sulfur Boiler - firetube sulfur boiler with forced hot water economizer to vaporize sulfur for heat transfer medium to regenerator. Total heat release 80 MM BTU/HR. Economizer heat recovery 20 MM BTU/HR. Thermal efficiency including economizer 75%. Hot water from economizer is used to temper reactor intercooler water system. Including auxiliaries.----	$ 70,000
15.	Incinerator and stack - 150 foot stack with gas fired incinerator designed to release purified smelter gas stream and sulfur boiler flue gas at a temperature of 800°F. ------------------	$150,000
	Total Purchased Equipment ----------------------------------	$2,030,000
	Total Fixed Capital --	$6,070,000

EXHIBIT 9-18

PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: H_2S (Low Temp. Claus) CASE NO. 1a

PROCESS: Direct Hot Gas Cat. Regen. GAS NO. 2B

Fixed Capital, Battery Limits Basis $ 6,895,000

Sulfur Production 196 NTD x 330 Days = 65,000 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	12.7 MCF	$.30 Avg.	$ 3.81	$ 248.0
1.	Total Raw Materials	-	-	$ 3.81	$ 248.0
2.	Direct Labor	18,000MH	$3.50	.97	63.0
3.	Supervision	1.	$10,000	.15	10.0
4.	Maintenance	5% FC	-	5.30	345.0
5.	Plant Supplies	10%(L2+L3)	-	.11	7.3
	Utilities:				
6. a)	Electricity	683 KWH	$.01	6.83	443.0
b)	Cooling Water	4.4 MG	.05	.22	14.3
c)	Boiler Feed Water	0.72 MG	.25	.18	11.7
d)	Fuel Gas	23.3 MCF	.30	7.00	455.0
7.	Total Utilities	-	-	14.23	924.0
8.	Other Direct (Catalyst)	77 Lbs.	.13	10.00	650.0
9.	TOTAL DIRECT	-	-	$ 34.57	$ 2,247.3
10.	Payroll Burden	30% L2	-	.29	18.9
11.	Plant Overhead	70%(L2+L3)	-	.79	51.1
12.	Pack & Ship	1000 MH	3.90	.06	3.9
13.	Waste Disposal	1000 MH	3.90	.06	3.9
14.	Other Indirect	-	-	-	-
15.	TOTAL INDIRECT	-	-	$ 1.20	$ 77.8
16.	Depreciation	10% FC	-	10.60	689.5
17.	Taxes	2% FC	-	2.12	137.9
18.	Insurance	1% FC	-	1.06	69.0
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$ 13.78	$ 896.4
21.	TOTAL OPERATING	-	-	$ 49.55	$ 3,221.5
22.	Credits None			-	-
23.	NET OPERATING	-	-	$ 49.55	$ 3,221.5

EXHIBIT 9-19

LOW TEMPERATURE CLAUS
H_2S REDUCTION OF SO_2
2B SMELTER GAS

CASE 1a (ALTERNATE)

ALTERNATE REGENERATOR SCHEME (Per Exhibit 5):

9A.	Direct hot inert gas moving bed regenerator, insulated 20 ft. diameter by 60 feet high with 37 ft. bed depth complete with gas distribution internals, 300 tons catalyst inventory and catalyst handling system, carbon steel construction.	$160,000
14A.	Inert gas heater - direct gas fired heater with economizer 61 MM BTU/hr. net heat release, 100 MM BTU/hr. gross heat release. Economizer forced hot water used to temper Claus converter cooling system 20 MM BTU/hr.	$150,000
12A.	Sulfur Condensers - as for Case I except 75,000 SCFM gas flow with 31 MM BTU/hr. transfer.	$250,000
13A.	Regenerator Circulation Blower - centrifugal two stage blower of cast iron casing construction suitable for service on nitrogen or inert gas only with ΔP of 10 PSI complete with 5000 HP motor.	$300,000

Substitution of costs of items 9A, 12A, 13A and 14A above (total $860,000) for items 9, 11, 12, 13 and 14 (total $585,000) will increase total purchased equipment cost from $2,030,000 to $2,305,000 with proportionate increase in plant fixed capital cost. Increased utilities cost will also result. Also weighing against this design is the expected increased agglomerization tendency of the bed.

Total Fixed Capital	$6,895,000

EXHIBIT 9-20
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: H_2S (Low Temp. Claus) CASE NO. 1b
PROCESS: Separate Reformer & H_2S Gen. GAS NO. 2B
Fixed Capital, Battery Limits Basis $ 6,070,000
Sulfur Production 196 NTD x 330 Days = 65,000 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	11.7 MCF	$.30 Avg.	$ 3.51	$ 228.0
1.	Total Raw Materials	-	-	$ 3.51	$ 228.0
2.	Direct Labor	18,000 MH	$3.50	.97	63.0
3.	Supervision	1.	$10,000	.15	10.0
4.	Maintenance	5% FC	-	4.67	303.5
5.	Plant Supplies	10%(L2+L3)	-	.11	7.3
	Utilities:				
6. a)	Electricity	230 KWH	$.01	2.30	149.0
b)	Cooling Water	3.4 MG	.05	.17	11.0
c)	Boiler Feed Water	0.72 MG	.25	.18	11.7
d)	Fuel Gas	23.3 MCF	.30	7.00	455.0
7.	Total Utilities	-	-	9.65	626.7
8.	Other Direct (Catalyst)	74 Lbs.	.13	9.60	624.0
9.	TOTAL DIRECT	-	-	$ 28.66	$ 1,862.5
10.	Payroll Burden	30% L2	-	.29	18.9
11.	Plant Overhead	70%(L2+L3)	-	.79	51.1
12.	Pack & Ship	1000 MH	3.90	.06	3.9
13.	Waste Disposal	1000 MH	3.90	.06	3.9
14.	Other Indirect	-	-	-	-
15.	TOTAL INDIRECT	-	-	$ 1.20	$ 77.8
16.	Depreciation	10% FC	-	9.35	607.0
17.	Taxes	2% FC	-	1.87	121.4
18.	Insurance	1% FC	-	.94	60.7
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$ 12.16	$ 789.1
21.	TOTAL OPERATING	-	-	$ 42.02	$ 2,729.4
22.	Credits None			-	-
23.	NET OPERATING	-	-	$ 42.02	$ 2,729.4

EXHIBIT 9-21

LOW TEMPERATURE CLAUS
H_2S REDUCTION OF SO_2
2B SMELTER GAS
CASE 1b
(SEPARATE H_2S GENERATION)

SUMMATION OF MAJOR EQUIPMENT

1. Methane Reformer - direct gas fired box type reforming furnace designed to reform 1600 to 2000 SCFM methane at 1570°F., 150 PSIG outlet incorporating charge preheat coils and steam generation coils in convection section. Thermal efficiency 85% including economizer coils. Steam generation 18000 lbs. per hr. at 250 PSIG. Complete with twin carbon bed desulfurizers.-- $260,000

   (Price courtesy of the Selas Corporation)

2. Hydrogen Sulfide Synthesis Converter - comprising a 5 ft. diameter x 12 ft. high packed section for gas cooling and reaction and a 10 ft. diameter by 8 ft. high reaction section having two catalyst sections 2 ft. deep. Complete with 8 tons of catalyst (lasts indef. like Claus Cat., F-1 Alumina)---- $ 35,000

3. Hydrogen Sulfide Reactor Intercooler - 150 PSIG steam generator used for interstage cooling of reaction stream, carbon steel construction.--$ 10,000

4. Gas Cooler - water cooled shell and tube carbon steel heat exchanger suitable for H_2S service, 15.0 MM BTU/hr.-------- $ 25,000

5. Liquid Sulfur Cooler - 150 PSIG steam generator used for cooling circulating sulfur stream. Carbon steel construction. 8000 lb./hr. steam generation.--------------------------- $ 8,000

6. Sulfur Pump Tank - carbon steel sulfur vertical hold tank with internal steam coils, 50 tons capacity, insulated.--------- $ 5,000

7. Sulfur circulation pump - 120 GPM steam jacketed pump complete with 10 HP motor and variable speed drive.---------------- $ 2,000

8. Low Temperature Claus Converter - moving catalyst bed design per exhibit 3. Chamber is steel, 20 ft. diameter x 50 ft. high straight side having three catalyst zones with internal downcomers and gas distribution system. Approximately 90% of gas flow is assumed to bypass through each interzone heat exchanger to control bed temperature rise. H_2S is added at several points to prevent excessive temperature rise in bed. Cooling medium for interzone gas coolers is tempered cooling water. Tube wall temperature is held at 250°F. minimum to minimize sulfur deposition on heat exchange surfaces. Including intercoolers, catalyst and auxiliaries.------------------- $730,000

EXH. 9-21 (Cont.)

9.	Catalyst Regenerator - indirect heated moving bed design comprising 22 ft. diameter x 50 ft. high carbon steel shell with 300 - 6" tubes 27' long through which catalyst passes countercurrent to an inert or reducing circulation gas stream. Catalyst is distributed above the upper tube sheet with a rake or similar mechanical distributor. Heat for sulfur stripping is provided by condensing sulfur vapor on external tube surfaces. Including catalyst and auxiliaries.	$320,000
10.	Catalyst Cooler - Carbon steel 15 ft. diameter by 20 ft. high shell with air distribution/catalyst support grates. Tempered air cools 100 NT per hour catalyst from 790°F. to 260°F. Complete with catalyst, catalyst handling equipment, air tempering coils, air blower with 100 HP motor and auxiliaries.	$ 80,000
11.	Regenerator Gas Exchanger - shell and tube carbon steel heat exchanger designed to reheat circulating stripping gas stream to 700°F. by exchange against regenerator overhead stream	$ 5,000
12.	Sulfur Coolers and Condensers - 35 PSIG steam generators designed to condense and recover 600 NT/day sulfur from regenerator gas stripping system. Carbon steel construction.	$100,000
13.	Regenerator Circulation Blower - centrifugal single stage blower of forged steel casing construction suitable for service on H_2S or nitrogen with ΔP of 8 PSI complete with 400HP motor.	$ 90,000
14.	Sulfur Boiler - firetube sulfur boiler with forced hot water economizer to vaporize sulfur for heat transfer medium to regenerator. Total heat release 80 MM BTU/hr. Economizer heat recovery 20 MM BTU/hr. Thermal efficiency including economizer 75%. Hot water from economizer is used to temper reactor intercooler water system. Including auxiliaries.	$ 70,000
15.	Incinerator and stack - 150 foot stack with gas fired incinerator designed to release purified smelter gas stream and sulfur boiler flue gas at a temperature of 800°F.	$150,000
16.	Smelter Gas Compressor - single stage centrifugal compressor to handle 70,000 SCFM at an inlet temperature of 77°F. with a ΔP of 3 PSI. Equipped with a 1500 HP motor.	$ 40,000
Total Purchased Equipment.		$2,030,000
Total Fixed Capital.		$6,070,000

EXHIBIT 9-23
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: H_2S (Low Temp. Claus) CASE NO. 1c

PROCESS: Water Injection Intercoolers GAS NO. 2B

Fixed Capital, Battery Limits Basis $4,440,000

Sulfur Production 196 NTD x 330 Days = 65,000 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	11.7 MCF	$.30 Avg.	$ 3.51	$ 228.0
1.	Total Raw Materials	-	-	$ 3.51	$ 228.0
2.	Direct Labor	18,000 MH	$3.50	.97	63.0
3.	Supervision	1.	$10,000	.15	10.0
4.	Maintenance	5% FC	-	3.42	222.0
5.	Plant Supplies	10%(L2+L3)	-	.11	7.3
	Utilities:				
6. a)	Electricity	230 KWH	$.01	2.30	149.5
b)	Cooling Water	1.1 MG	.05	.06	3.9
c)	Boiler Feed Water	0.72 MG	.25	.18	11.7
d)	Fuel Gas	23.8 MCF	.30	7.14	463.0
7.	Total Utilities	-	-	9.68	628.1
8.	Other Direct (Catalyst)	74 Lbs.	.13	9.60	624.0
9.	TOTAL DIRECT	-	-	$ 27.44	$ 1,782.4
10.	Payroll Burden	30% L2	-	.29	18.9
11.	Plant Overhead	70%(L2+L3)	-	.79	51.1
12.	Pack & Ship	1000 MH	3.90	.06	3.9
13.	Waste Disposal	1000 MH	3.90	.06	3.9
14.	Other Indirect	-	-	-	-
15.	TOTAL INDIRECT	-	-	$ 1.20	$ 77.8
16.	Depreciation	10% FC	-	6.83	444.0
17.	Taxes	2% FC	-	1.37	88.8
18.	Insurance	1% FC	-	.68	44.4
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$ 8.88	$ 577.2
21.	TOTAL OPERATING	-	-	$ 37.52	$ 2,437.4
22.	Credits None			-	-
23.	NET OPERATING	-	-	$ 37.52	$ 2,437.4

EXHIBIT 9-24

LOW TEMPERATURE CLAUS
H_2S REDUCTION OF SO_2
2B SMELTER GAS
CASE IC
SUMMATION OF MAJOR EQUIPMENT

1.	Methane Reformer - direct gas fired box type reforming furnace designed to reform 1600 to 2000 SCFM methane at 1570°F., 150 PSIG outlet incorporating charge preheat coils and steam generation coils in convection section. Thermal efficiency 85% including economizer coils. Steam generation 18000 lb./hr. at 250 PSIG. Complete with twin carbon bed desulfurizers. (Price courtesy of the Selas Corporation)	$260,000
2.	Hydrogen Sulfide Synthesis Converter - comprising a 5 ft. diameter x 12 ft. high packed section for gas cooling and reaction and a 10 ft. diameter by 8 ft. high reaction section having two catalyst sections 2 ft. deep. Complete with 8 tons of catalyst.	$ 35,000
3.	Hydrogen Sulfide Reactor Intercooler - 150 PSIG steam generator used for interstage cooling of reaction stream, carbon steel construction.	$ 10,000
4.	Gas Cooler - water cooled shell and tube carbon steel heat exchanger suitable for H_2S service, 34.0 MM BTU/hr.	$ 55,000
5.	Liquid Sulfur Cooler - 150 PSIG steam generator used for cooling circulating sulfur stream. Carbon steel construction. 8000 lb./hr. steam generation.	$ 8,000
6.	Sulfur Pump Tank - carbon steel sulfur vertical hold tank with internal steam coils, 50 tons capacity, insulated.	$ 5,000
7.	Sulfur circulation pump - 120 GPM steam jacketed pump complete with 10 HP motor and variable speed drive.	$ 2,000
8.	Low Temperature Claus Converter - moving catalyst bed design per exhibit 3. Chamber is steel, 21 ft. diameter x 50 ft. high straight side having three catalyst zones with internal downcomers and gas distribution system. Approximately 90% of gas flow is assumed to bypass through each interzone water injection chamber to control bed temperature rise. H_2S is added at several points to prevent excessive temperature rise in bed. Including injection chambers, catalyst and auxiliaries.	$150,000

9. Catalyst Regenerator - indirect heated moving bed design comprising 22 ft. diameter x 50 ft. high carbon steel shell with 380 - 6" tubes 27' long through which catalyst passes countercurrent to an inert or reducing circulation gas stream. Catalyst is distributed above the upper tube sheet with a rake or similar mechanical distributor. Heat for sulfur stripping is provided by condensing sulfur vapor on external tube surfaces. Including catalyst and auxiliaries.-------- $320,000

10. Catalyst Cooler - Carbon steel 15 ft. diameter by 20 ft. high shell with air distribution/catalyst support grates. Tempered air cools 100 NT per hour catalyst from 790°F. to 260°F. Complete with catalyst, catalyst handling equipment, air tempering coils, air blower with 100 HP motor and auxiliaries.-- $ 80,000

11. Regenerator Gas Exchanger - shell and tube carbon steel heat exchanger designed to reheat circulating stripping gas stream to 700°F. by exchange against regenerator overhead stream. $ 5,000

12. Sulfur Coolers and Condensers - 35 PSIG steam generators designed to condense and recover 600 NT/day sulfur from regenerator gas stripping system. Carbon steel construction.-- $100,000

13. Regenerator Circulation Blower - centrifugal single stage blower of forged steel casing construction suitable for service on H_2S or nitrogen with ΔP of 8 PSI complete with 400HP motor.-- $ 90,000

14. Sulfur Boiler - firetube sulfur boiler with 150 PSIG steam economizer to vaporize sulfur for heat transfer medium to regenerator. Total heat release 80 MM BTU/hr. Economizer heat recovery 20 MMBTU/hr. Thermal efficiency including economizer 75%. Including auxiliaries.------------------- $ 70,000

15. Incinerator and stack - 150 foot stack with gas fired incinerator designed to release purified smelter gas stream and sulfur boiler flue gas at a temperature of 800°F.------ $150,000

16. Smelter Gas Compressor - single stage centrifugal compressor to handle 70,000 SCFM at an inlet temperature of 77°F. with a ΔP of 3 PSI. Equipped with a 1500 HP motor.------------- $100,000

17. Sulfur Storage Tank - vertical steel tank 2,000 ton capacity with steam coils and external insulation.----------------- $ 40,000

Total Purchased Equipment ---------------------------------- $1,480,000

Total Fixed Capital -- $4,440,000

EXHIBIT 9-25
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: H_2S (Low Temp. Claus) CASE NO. 1d
PROCESS: Adiabatic Reactor-Vent Recycle CAS NO. 2B

Fixed Capital, Battery Limits Basis $5,060,000
Sulfur Production 196 NTD x 330 Days = 65,000 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	11.7 MCF	$.30 Avg.	$ 3.51	$ 228.0
1.	Total Raw Materials	-	-	$ 3.51	$ 228.0
2.	Direct Labor	18,000 MH	$3.50	.97	63.0
3.	Supervision	1.	$10,000	.15	10.0
4.	Maintenance	5% FC	-	3.90	253.0
5.	Plant Supplies	10%(L2+L3)	-	.11	7.3
	Utilities:				
6. a)	Electricity	420 KWH	$.01	4.20	273.0
b)	Cooling Water	4.4 MG	.05	.22	14.3
c)	Boiler Feed Water	0.72 MG	.25	.18	11.7
d)	Fuel Gas	23.3 MCF	.30	7.00	455.0
7.	Total Utilities	-	-	11.60	753.0
8.	Other Direct (Catalyst)	74 Lbs.	.13	9.60	624.0
9.	TOTAL DIRECT	-	-	$ 29.84	$ 1,938.3
10.	Payroll Burden	30% L2	-	.29	18.9
11.	Plant Overhead	70%(L2+L3)	-	.79	51.1
12.	Pack & Ship	1000 MH	3.90	.06	3.9
13.	Waste Disposal	1000 MH	3.90	.06	3.9
14.	Other Indirect	-	-	-	-
15.	TOTAL INDIRECT	-	-	$ 1.20	$ 77.8
16.	Depreciation	10% FC	-	7.78	506.0
17.	Taxes	2% FC	-	1.56	101.2
18.	Insurance	1% FC	-	.78	50.6
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$ 10.12	$ 657.8
21.	TOTAL OPERATING	-	-	$ 41.16	$ 2,673.9
22.	Credits None			-	-
23.	NET OPERATING	-	-	$ 41.16	$ 2,673.9

EXHIBIT 9-26

LOW TEMPERATURE CLAUS
H_2S REDUCTION OF SO_2
2B SMELTER GAS
CASE 1d
SUMMATION OF MAJOR EQUIPMENT

1.	Methane Reformer - direct gas fired box type reforming furnace designed to reform 1600 to 2000 SCFM methane at 1570°F., 150 PSIG outlet incorporating charge preheat coils and steam generation coils in convection section. Thermal efficiency 85% including economizer coils. Steam generation 18000 lb./hr. at 250 PSIG. Complete with twin carbon bed desulfurizers.--- (Price courtesy of the Selas Corporation)	$260,000
2.	Hydrogen Sulfide Synthesis Converter - comprising a 5 ft. diameter x 12 ft. high packed section for gas cooling and reaction and a 10 ft. diameter by 8 ft. high reaction section having two catalyst sections 2 ft. deep. Complete with 8 tons of catalyst.--	$ 35,000
3.	Hydrogen Sulfide Reactor Intercooler - 150 PSIG steam generator used for interstage cooling of reaction stream, carbon steel construction.--	$ 10,000
4.	Gas Cooler - water cooled shell and tube carbon steel heat exchanger suitable for H_2S service, 15.0 MM BTU/hr.----------	$ 25,000
5.	Liquid Sulfur Cooler - 150 PSIG steam generator used for cooling circulating sulfur stream. Carbon steel construction. 8000 lb./hr. steam generation.------------------------------	$ 8,000
6.	Sulfur Pump Tank - carbon steel sulfur vertical hold tank with internal steam coils, 50 tons capacity, insulated.------	$ 5,000
7.	Sulfur circulation pump - 120 GPM steam jacketed pump complete with 10 HP motor and variable speed drive.------------------	$ 2,000
8.	Low Temperature Claus Converter - moving catalyst bed design per exhibit 3. Chamber is steel, 28.5 ft. diameter x 20 ft. high straight side having one catalyst zone 6 ft. deep. Including catalyst and auxiliaries.-------------------------	$200,000
9.	Catalyst Regenerator - indirect heated moving bed design comprising 22 ft. diameter x 50 ft. high carbon steel shell with 380 - 6" tubes 27' long through which catalyst passes countercurrent to an inert or reducing circulation gas stream. Catalyst is distributed above the upper tube sheet with a rake or similar mechanical distributor. Heat for sulfur stripping is provided by condensing sulfur vapor on external tube surfaces. Including catalyst and auxiliaries.----------	$300,000

10.	Catalyst Cooler - Carbon steel 15 ft. diameter by 20 ft. high shell with air distribution/catalyst support grates. Tempered air cools 100 NT per hour catalyst from 790°F. to 260°F. Complete with catalyst, catalyst handling equipment, air tempering coils, air blower with 100 HP motor and auxiliaries.	$ 80,000
11.	Regenerator Gas Exchanger - shell and tube carbon steel heat exchanger designed to reheat circulating stripping gas stream to 700°F. by exchange against regenerator overhead stream.	$ 5,000
12.	Sulfur Coolers and Condensers - 35 PSIG steam generators designed to condense and recover 600 NT/day sulfur from regenerator gas stripping system. Carbon steel construction.	$ 100,000
13.	Regenerator Circulation Blower - centrifugal single stage blower of forged steel casing construction suitable for service on H_2S or nitrogen with ΔP of 8 PSI complete with 400HP motor.	$ 90,000
14.	Sulfur Boiler - firetube sulfur boiler with 150 PSIG steam economizer to vaporize sulfur for heat transfer medium to regenerator. Total heat release 80 MM BTU/hr. Economizer heat recovery 20 MM BTU/hr. Thermal efficiency including economizer 75%. Including auxiliaries.	$ 70,000
15.	Incinerator and stack - 150 foot stack with gas fired incinerator designed to release purified smelter gas stream and sulfur boiler flue gas at a temperature of 800°F.	$150,000
16.	Smelter Gas Compressor - single stage centrifugal compressor to handle 70,000 SCFM at an inlet temperature of 77°F. with a ΔP of 3 PSI. Equipped with a 1500 HP motor.	$100,000
17.	Sulfur Storage Tank - vertical steel tank 2,000 ton capacity with steam coils and external insulation.	$ 40,000
18.	Recycle Gas Blower - single stage centrifugal compressor to handle 70,000 SCFM at an inlet temperature of 350°F. and a ΔP of 4 PSI complete with 2000 HP motor.	$130,000
19.	Recycle Gas Cooler - carbon steel shell and tube exchanger suitable for cooling 70,000 SCFM gas from 350°F. to 212°F. using water as coolant.	$ 80,000
	Total Purchased Equipment	$1,690,000
	Total Fixed Capital	$5,060,000

EXHIBIT 9-27R
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: H_2S CASE NO. 2R
PROCESS: Low Temp. Claus GAS NO. 3

Fixed Capital, Battery Limits Basis $ 16,800,000
Sulfur Production 385 NTD x 330 Days = 127,000 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	12.4 MCF	$.30 Avg.	$ 3.72	$ 472.0
1.	Total Raw Materials	-	-	$ 3.72	$ 472.0
2.	Direct Labor	27,000 MH	$3.50	.74	94.5
3.	Supervision	1.	$10,000	.12	15.0
4.	Maintenance	5% FC	-	6.61	840.0
5.	Plant Supplies	10%(L2+L3)	-	.09	11.0
	Utilities:				
6. a)	Electricity	2,580 KWH	$.006	15.45	1,962.0
b)	Cooling Water	0.46 MG	.05	.02	2.5
c)	Boiler Feed Water	1.40 MG	.25	.35	44.5
d)	Fuel Gas	18.7 MCF	.30	5.62	715.0
7.	Total Utilities	-	-	21.44	2,724.0
8.	Other Direct (Catalyst)	77 Lbs.	.13	10.00	1,270.0
9.	TOTAL DIRECT	-	-	$ 42.72	$ 5,426.5
10.	Payroll Burden	30% L2	-	.22	28.4
11.	Plant Overhead	70%(L2+L3)	-	.61	76.7
12.	Pack & Ship	1000 MH	3.90	.03	3.9
13.	Waste Disposal	1000 MH	3.90	.03	3.9
14.	Other Indirect	-	-		
15.	TOTAL INDIRECT	-	-	$ .89	$ 112.9
16.	Depreciation	10% FC	-	13.22	1,680.0
17.	Taxes	2% FC	-	2.64	336.0
18.	Insurance	1% FC	-	1.32	168.0
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$ 17.18	$ 2,184.0
21.	TOTAL OPERATING	-	-	$ 60.79	$ 7,723.4
22.	Credits	8.6 M Lb.	.25	2.15	273.0
23.	NET OPERATING	-	-	$ 58.64	$ 7,450.4

EXHIBIT 9-28

LOW TEMPERATURE CLAUS
H_2S REDUCTION OF SO_2
#3 STACK GAS
CASE 2
SUMMATION OF MAJOR EQUIPMENT

1.	Methane Reformer - direct gas fired box type reforming furnace designed to reform 3600 SCFM methane at 1570°F., 150 PSIG outlet incorporating charge preheat coils and steam generation coils in convection section. Thermal efficiency 85% including economizer coils. Steam generation 40,000 lb./hr. at 250 PSIG. Complete with twin carbon bed desulfurizers.----------------	$425,000
2.	Hydrogen Sulfide Synthesis Converter - comprising a 7 ft. diameter x 12 ft. high packed section for gas cooling and reaction and a 14 ft. diameter by 8 ft. high reaction section having two catalyst beds 2 ft. deep. Complete with catalyst.----------	$ 60,000
3.	Hydrogen Sulfide Reactor Intercooler - 150 PSIG steam generator used for interstage cooling of reaction stream, carbon steel construction, 22,000 lb./hr. steam generation.--------------	$ 25,000
4.	Gas Cooler - water cooled shell and tube carbon steel heat exchanger suitable for H_2S service 34.0 MM BTU/hr.----------	$ 55,000
5.	Liquid Sulfur Cooler - 150 PSIG steam generator used for cooling circulating sulfur stream. Carbon steel construction. 18,000 lb./hr. steam generation.----------------------------------	$ 18,000
6.	Sulfur Pump Tank - carbon steel vertical sulfur hold tank with internal steam coils, 100 tons capacity, insulated.---------	$ 8,000
7.	Sulfur Circulation Pump - 250GPM steam jacketed pump complete with 30 HP motor and variable speed drive.------------------	$ 3,000
8.	Flue Gas Damper and Duct System - carbon steel construction, included as major capital equipment.------------------------	$100,000
9.	Low Temperature Claus Converter - Rectangular 40 ft. by 100 ft. chamber 80 ft. high complete with four levels of moving grates. Each grate 4 ft. by 100 feet, 10 grates per level. Complete with catalyst bed 2.5 feet deep on all grates (total 1000 tons), all internals, baffles supports, drives and motors, and catalyst handling system.--	$3,157,000
10.	Gas Booster Fan - single stage blower for 2.4 million SCFM at 255°F. with ΔP = 30 inches W.C. carbon steel construction complete with 25,000 HP motor.------------------------------	$400,000

EXH. 9-28 (Cont.)

11.	Catalyst Regenerator - indirect heated moving bed design comprising 33 ft. diameter x 50 ft. high carbon steel shell with 820 - 6" tubes 37 ft. long through which catalyst passes countercurrent to an inert or reducing circulation gas stream. Catalyst is distributed above the upper tubesheet with a rake or similar mechanical distributor. Heat for sulfur stripping is provided by condensing sulfur vapor on external tube surfaces. Including catalyst and auxiliaries.	$625,000
12.	Catalyst Cooler - Carbon steel 22 ft. diameter by 30 ft. high shell with air distribution/catalyst support gratings. Tempered air cools 220 NT per hour catalyst from 790°F. to 260°F. Complete with catalyst. Catalyst handling equipment, air tempering coils, air blower with 200 HP motor, and auxiliaries.	$130,000
13.	Regenerator Circulation Blower - centrifugal single stage blower of forged steel casing construction suitable for service on H_2S or nitrogen with ΔP of 8 PSI complete with 800 HP motor	$160,000
14.	Sulfur Boiler - firetube sulfur boiler with 150 PSIG steam economizer to vaporize sulfur for heat transfer medium to regenerator. Total heat release 172 MM BTU/hr. Economizer heat recovery 43 MM BTU/hr. steam generation 43,000 lb./hr. Thermal efficiency including economizer 75%. Including auxiliaries.	$120,000
15.	Regenerator Gas Exchanger - shell and tube carbon steel heat exchanger designed to reheat circulating stripping gas stream to 700°F. by exchange against regenerator overhead stream	$ 10,000
16.	Sulfur Coolers and Condensers - 35 PSIG steam generators designed to condense and recover 1300 NT/day sulfur from regenerator gas stripping system. Carbon steel construction	$170,000
17.	Sulfur Storage Tank - vertical steel tank 2,000 ton capacity with steam coils and external insulation.	$ 40,000
	Total Purchased Equipment	$5,506,000
	Total Fixed Capital	$16,500,000

(See Exhibits 9-27R and 4-2R)

EXHIBIT 9-28R

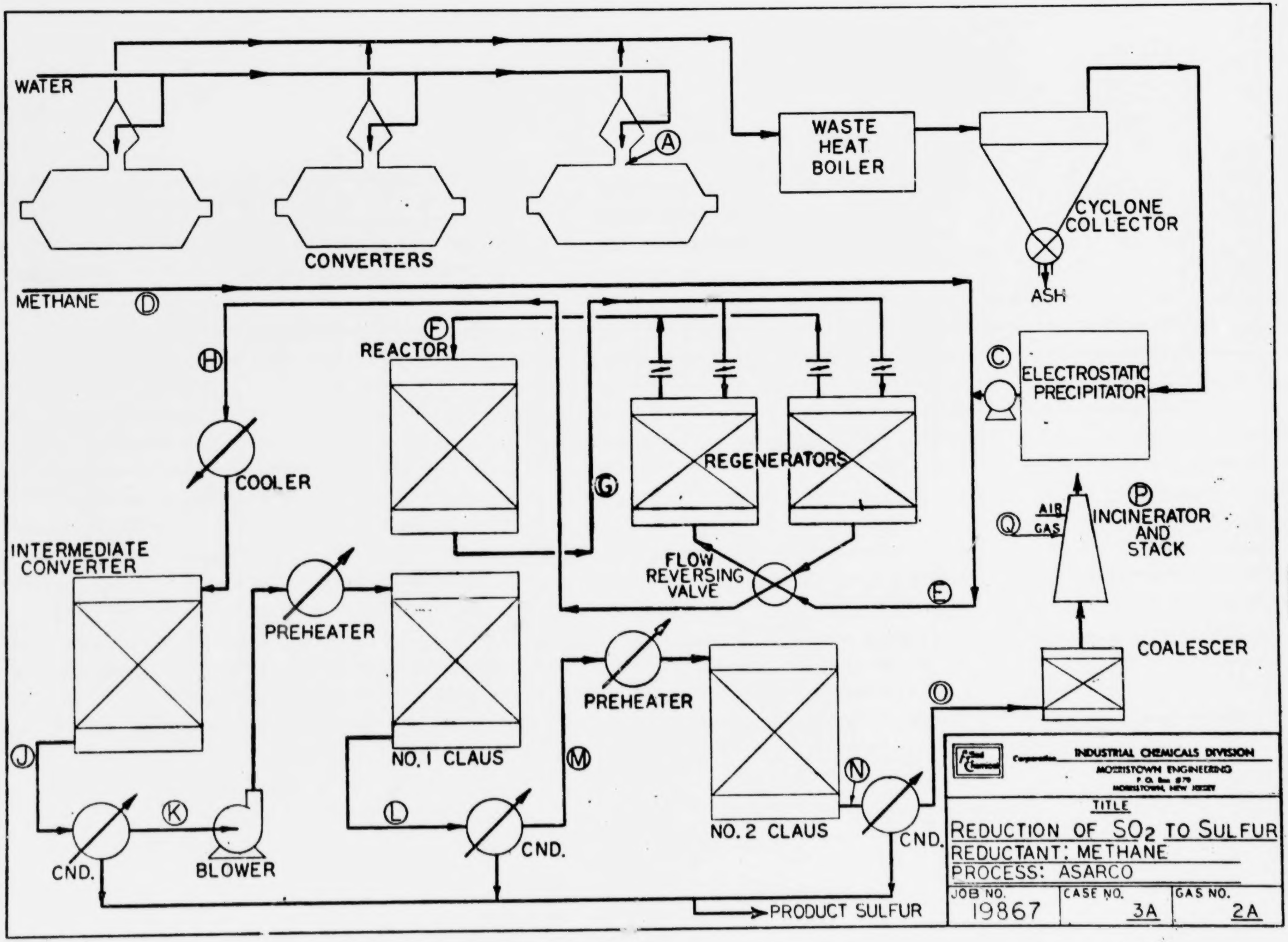
Change Exh. 9-28 to read:

CASE 2R (REVISED)

Item 9	Low Temp. Claus Converter per level. Each catalyst bed to be 2.5 feet deep on all grates, including all internals, baffles supports, drives and motors plus catalyst handling system, but excluding catalyst cost ------------------------	$2,950,000
Item 10	Gas Booster Fan - single stage blower for 2.4 million SCFM at 255°F with sp = 60 inches W.C. carbon steel construction, complete with 50,000 HP motor -------------------------------	$ 650,000

Total purchased equipment ------------------------------	$5,506,000
Fixed Capital less Catalyst ----------------------------	$16,550,000
Initial Catalyst Charge, 1000 tons ---------------------	$ 250,000
Total Fixed Capital ------------------------------------	$16,800,000

EXHIBIT 10-1


WATER
CONVERTERS
A
WASTE HEAT BOILER
CYCLONE COLLECTOR
ASH
METHANE
D
H
F
REACTOR
C
ELECTROSTATIC PRECIPITATOR
COOLER
G
REGENERATORS
P
INCINERATOR AND STACK
AIR
GAS
Q
INTERMEDIATE CONVERTER
FLOW REVERSING VALVE
E
PREHEATER
PREHEATER
COALESCER
O
J
NO. 1 CLAUS
M
NO. 2 CLAUS
N
K
L
CND.
BLOWER
CND.
CND.
PRODUCT SULFUR
Allied Chemical Corporation
INDUSTRIAL CHEMICALS DIVISION
MORRISTOWN ENGINEERING
P. O. Box 879
MORRISTOWN, NEW JERSEY
TITLE
REDUCTION OF SO2 TO SULFUR
REDUCTANT: METHANE
PROCESS: ASARCO
JOB NO. 19867
CASE NO. 3A
GAS NO. 2A

DRAWING NO. 37193

Exhibit 10-2

METHANE REDUCTION OF SMELTER GAS

CASE 3a — 20,000 SCFM — ASARCO PROCESS

GAS 2a

Process Points	A	B	C	D	E	F	G	H	J	K	L	M	N	O	P	Q
Temp. °F	2300	685	658	60	588	2323	2282	934/734	910	260	491	260	412	250	1000	60
Vol. MSCFM	20.0	21.6	21.6	1.70	23.2	23.2	25.0	25.0	24.7	23.6	23.4	23.4	23.4	23.3	28.7	4.70
Vol. MSCFM at T°F	112.0	50.2	52.5	1.80	49.4	131.0	139.5	60.6	-	34.6	43.3	34.6	40.5	33.7	85.1	5.00
Vol. % CH_4	-	-	-	100.0	7.33	6.82	-	-	-	-	-	-	-	-	-	100.0
% SO_2	15.8	14.7	14.7	-	13.6	13.6	2.39	2.39	1.27	1.33	0.25	0.25	0.12	0.12	0.57	-
% H_2O	.002	7.21	7.21	-	6.63	7.72	17.0	17.0	17.6	18.3	20.4	20.7	21.0	21.0	20.45	-
% H_2S	-	-	-	-	-	-	1.20	1.20	2.54	2.65	0.50	0.506	0.234	0.235	-	-
% CO_2	0.0	-	-	-	-	0.52	4.84	4.84	6.89	7.18	7.16	7.28	7.27	7.26	7.51	-
% CO	-	-	-	-	-	-	1.92	-	-	-	-	-	-	-	-	-
% S_2	-	-	-	-	-	-	4.52	4.52	4.52	0.184	1.79	0.186	0.39	0.092	-	-
% H_2	-	-	-	-	-	-	1.63	1.63	-	-	-	-	-	-	-	-
% COS	-	-	-	-	-	-	0.05	0.05	-	-	-	-	-	-	-	-
% N_2	83.0	77.0	77.0	-	71.4	71.4	66.4	66.4	67.2	70.2	70.0	71.0	71.0	71.3	71.46	-
% O_2	1.2	1.12	1.12	-	1.03	0.00	-	-	-	-	-	-	-	-	-	-

POINT	SULFUR DISTRIBUTION		
W	From Intermed. Converter Cond.	136.4	NT/Day
X	From No. 1 Claus Cond.	48.67	"
Y	From No. 2 Claus Cond.	8.90	"
Z	Total (Incl. Coalescer)	194.0	"

OVERALL CONVERSION 99.1 %

SULFUR YIELD 96.0 %

METHANE FOR RED'N: 12.7 MSCF/TON S

(13.4 at 60°F)

For Fuel: 5.0 MCF/Ton S

Exhibit 10-3

METHANE REDUCTION OF SMELTER GAS

CASE 3b

GAS 2a

20,000 SCFM

WEST PROCESS

Process Points	A	B	C	D	E	F	G	H	J	K	L	M	N	O	P	Q
Temp. °F	2300	685	658	68	588	1533	1832	759	740	260	450	260	425	250	1000	600
Vol. MSCFM	20.0	21.6	21.6	1.70	23.2	23.2	24.8	24.8	24.7	23.6	23.8	23.4	23.4	23.3	28.7	0.634
Vol. MSCFM at T°F	112.0	50.2	52.5	1.82	49.4	94.0	115.6	61.5	60.3	34.5	44.1	34.2	42.1	33.7	85.2	0.672
Vol. % CH_4	-	-	-	100.	7.31	6.79	0.00	-	-	-	-	-	-	-	-	100
% SO_2	15.8	14.7	14.7	-	13.6	13.6	1.81	1.81	1.36	1.42	0.24	0.24	0.16	0.16	0.54	-
% H_2O	.002	7.21	7.21	-	6.68	7.72	11.76	11.76	17.37	18.13	20.37	20.73	20.89	20.94	20.39	-
% H_2S	-	-	-	-	-	-	2.16	2.16	2.72	2.84	0.48	0.49	0.32	0.32	-	-
% CO_2	0.00	-	-	-	-	0.52	6.59	6.59	6.89	7.19	7.15	7.28	7.28	7.29	7.48	-
% CO	-	-	-	-	-	-	0.67	0.67	-	-	-	-	-	-	-	-
% S_2	-	-	-	-	-	-	4.79	4.79	4.37	0.18	1.94	0.19	0.31	0.09	-	-
% H_2	-	-	-	-	-	-	0.72	0.72	-	-	-	-	-	-	-	-
% COS	-	-	-	-	-	-	0.06	0.06	-	-	-	-	-	-	-	-
% N_2	83.0	77.0	77.0	-	71.3	71.4	71.45	71.45	67.3	70.24	69.83	71.07	71.04	71.20	71.60	-
% O_2	1.20	1.11	1.11	-	1.03	-	-	-	-	-	-	-	-	-	-	-

POINT	SULFUR DISTRIBUTION		
W	From Intermed. Converter Cond.	132.2	MT/Day
X	From No. 1 Claus Cond.	53.3	"
Y	From No. 2 Claus Cond.	6.6	"
Z	Total (Incl. Coalescer)	192.1	"

OVERALL CONVERSION 98.8 %

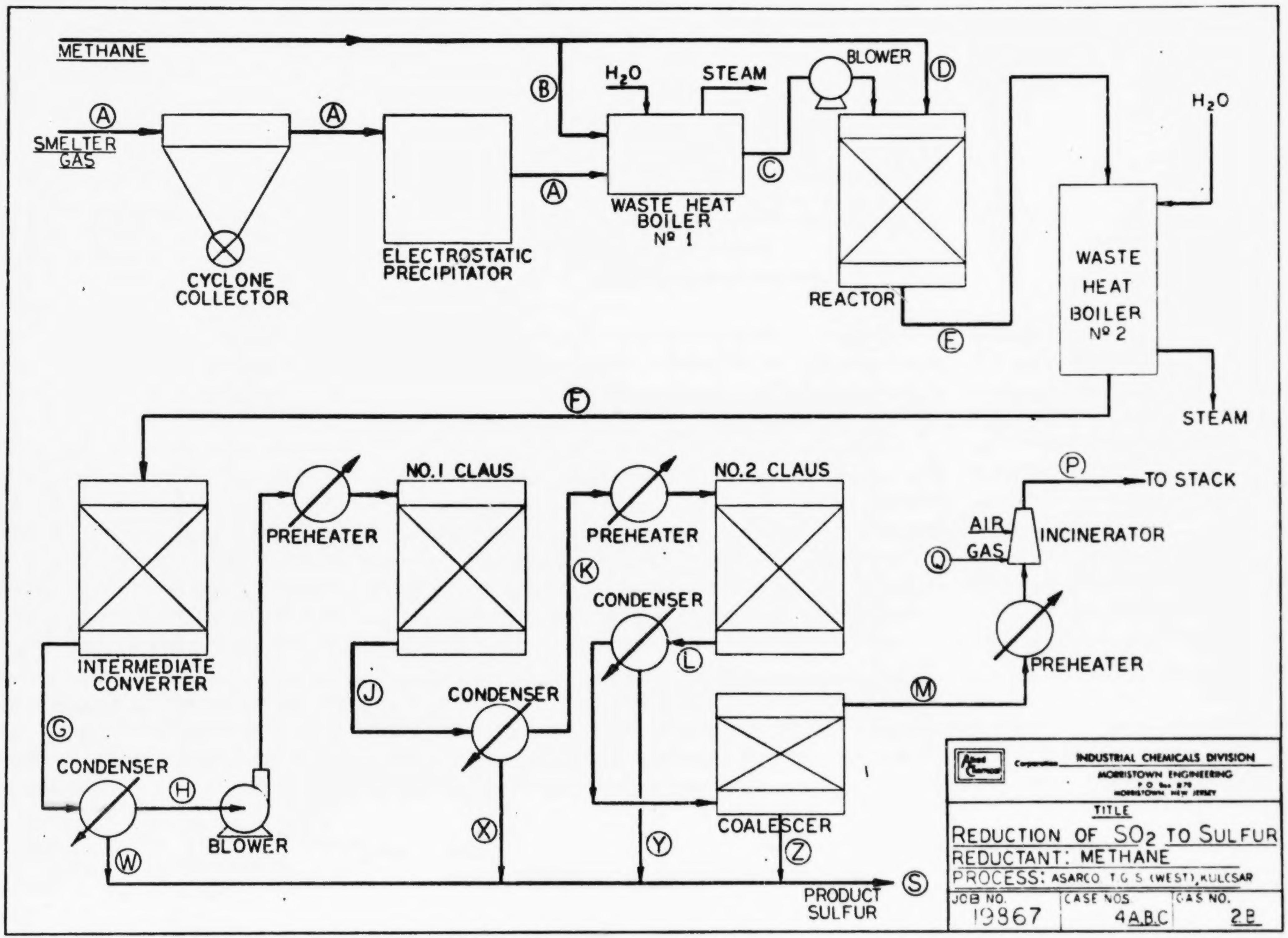
SULFUR YIELD 95.1 %

METHANE FOR RED'N: 12.7 MSCF/TON S

(13.4 at 60°F)

For Fuel: 5.0 MCF/Ton S

EXHIBIT 10-4


METHANE
Ⓐ
SMELTER GAS
CYCLONE COLLECTOR
Ⓐ
ELECTROSTATIC PRECIPITATOR
Ⓐ
Ⓑ
H_2O
STEAM
WASTE HEAT BOILER Nº 1
Ⓒ
BLOWER
Ⓓ
REACTOR
Ⓔ
H_2O
WASTE HEAT BOILER Nº 2
STEAM
Ⓕ
INTERMEDIATE CONVERTER
Ⓖ
CONDENSER
Ⓗ
BLOWER
Ⓦ
PREHEATER
NO.1 CLAUS
Ⓙ
CONDENSER
Ⓧ
PREHEATER
Ⓚ
NO.2 CLAUS
CONDENSER
Ⓛ
Ⓨ
COALESCER
Ⓩ
PRODUCT SULFUR
Ⓢ
Ⓜ
PREHEATER
Ⓠ
AIR
GAS
INCINERATOR
Ⓟ
TO STACK

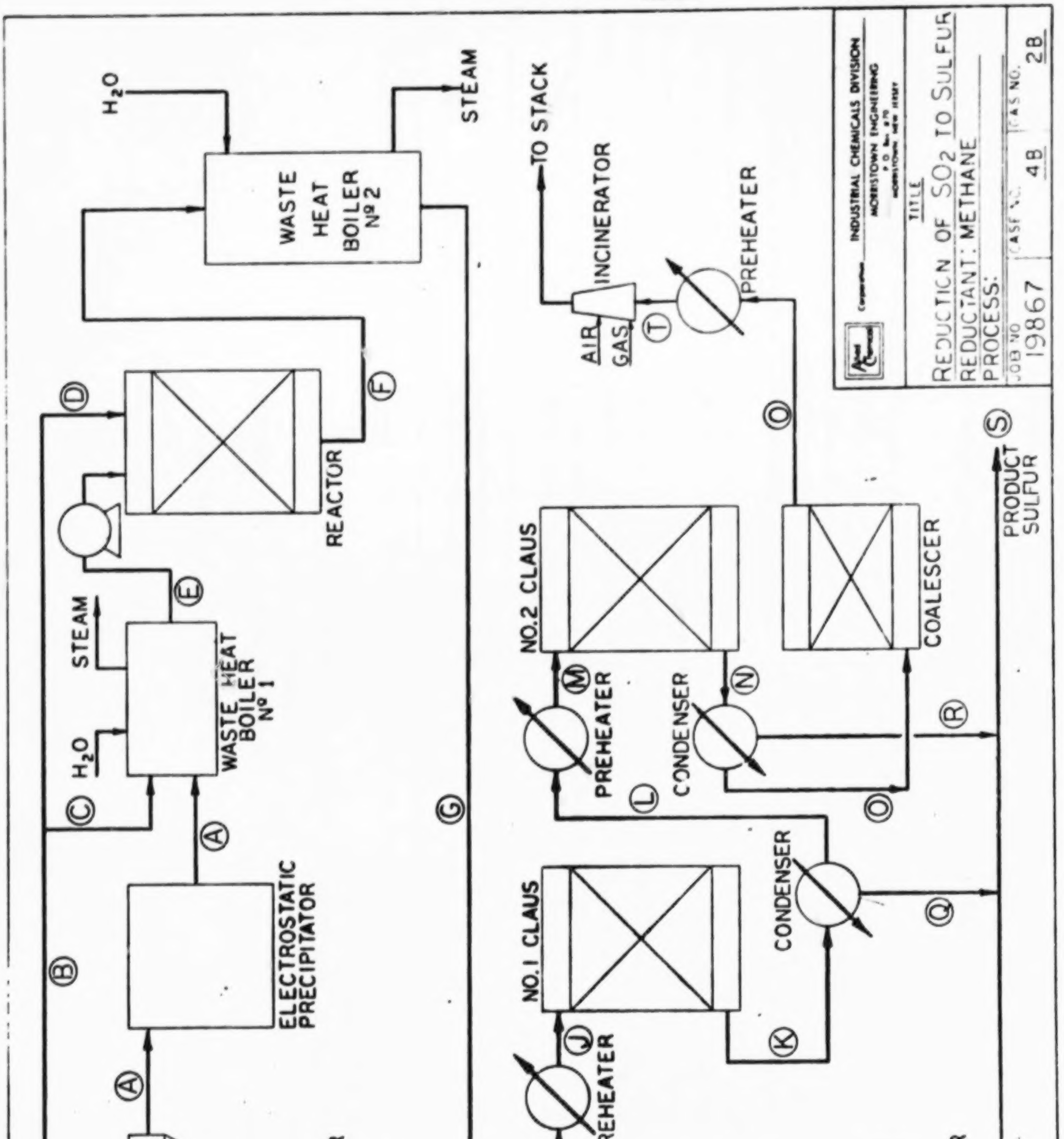
Corporation INDUSTRIAL CHEMICALS DIVISION
MORRISTOWN ENGINEERING
P.O. Box 879
MORRISTOWN, NEW JERSEY

TITLE
REDUCTION OF SO_2 TO SULFUR
REDUCTANT: METHANE
PROCESS: ASARCO T.G.S. (WEST), KULCSAR

JOB NO.	CASE NOS.	GAS NO.
19867	4 A,B,C	2B

DRAWING NO. 37194

EXHIBIT 10-4B


METHANE
SMELTER GAS
A
CYCLONE COLLECTOR
A
ELECTROSTATIC PRECIPITATOR
B
A
C
H2O
STEAM
WASTE HEAT BOILER Nº 1
E
D
REACTOR
F
WASTE HEAT BOILER Nº 2
H2O
STEAM
G
INTERMEDIATE CONVERTER
H
CONDENSER
I
P
BLOWER
PREHEATER
J
NO.1 CLAUS
K
CONDENSER
Q
L
PREHEATER
M
NO.2 CLAUS
CONDENSER
N
O
R
COALESCER
PRODUCT SULFUR
S
O
PREHEATER
T
AIR
GAS
INCINERATOR
TO STACK

Allied Chemical Corporation
INDUSTRIAL CHEMICALS DIVISION
MORRISTOWN ENGINEERING
P.O. Box 970
MORRISTOWN NEW JERSEY

TITLE
REDUCTION OF SO_2 TO SULFUR
REDUCTANT: METHANE
PROCESS:

JOB NO.	CASE NO.	FAS NO.
19867	4B	2B

DRAWING NO 37195

METHANE REDUCTION OF Smelter Gas

CASE 4a
GAS 2b

70,000 SCFM

ASARCO PROCESS

Process Points	A	B	C	D	E	F	G	H	J	K	L	M	N	O	P	Q	S
Temp. °F	440	60.	2457	60.	2282	734	894	260	505	260	405	250	-	-	1200	60	250
Vol. MSCFM	70.0	5.49	75.5	1.86	79.6	79.6	78.5	72.4	78.7	77.7	77.1	77.5	-	-	77.5	795	-
Vol. MSCFM at T°F	128.0	5.80	447.	1.97	443.	193.	216.	114.	154.	114.	137.	153.	-	-	230.	840	-
Vol. % CH_4	-	100.	-	100.	-	-	-	-	-	-	-	-	-	-	-	-	-
% SO_2	4.50	-	3.79	-	2.07	2.07	1.08	1.08	0.211	0.214	0.105	0.105	-	-	0.497	-	-
% H_2O	-	-	13.2	-	16.3	16.3	16.6	16.6	18.3	18.5	18.7	18.8	-	-	18.4	-	-
% H_2S	-	-	-	-	0.473	0.473	2.14	2.15	0.423	0.428	0.209	0.210	-	-	0.00	-	-
% CO_2	-	-	6.59	-	7.24	7.24	9.36	9.36	9.34	9.46	9.46	9.48	-	-	9.15	-	-
% CO	-	-	-	-	1.98	1.98	-	-	-	-	-	-	-	-	-	-	-
% S_2	-	-	-	-	0.697	0.697	0.414	0.197	1.48	0.199	0.363	0.100	-	-	0.00	-	100.
% H_2	-	-	-	-	1.68	1.68	0.002	-	-	-	-	-	-	-	-	-	-
% COS	-	-	-	-	0.018	0.018	0.004	-	-	-	-	-	-	-	-	-	-
% N_2	79.0	-	66.4	-	69.5	69.5	70.4	70.6	70.3	70.2	71.2	71.4	-	-	72.0	-	-
% O_2	16.5	-	0.693	-	-	-	-	-	-	-	-	-	-	-	-	-	-

POINT	SULFUR DISTRIBUTION		
W	From Intermed. Converter Cond.	21.9	NT/Day
X	From No. 1 Claus Cond.	130.4	"
Y	From No. 2 Claus Cond.	26.4	"
S	Total (Incl. Coalescer)	178.7	"

Includes CH_4 for H_2S Incineration & Preheaters

OVERALL CONVERSION 97.4 %

SULFUR YIELD 88.0 %

METHANE FOR RED'N: 15.0 MSCF/TON S
(15.8 at 60°F)

For Fuel: 46.5 MCF/Ton S

Exhibit 10-5

Exhibit 10-6

METHANE REDUCTION OF SMELTER GAS

CASE 4b

GAS 2b

70,000 SCFM

WEST PROCESS

Process Points	A	B	C	D	E	F	G	H	J	K	L	M	N	O	P	Q	S
Temp. °F	440	60	1854	60	1832	734	894	260	505	260	405	250	-	-	1200	60	250
Vol. MSCFM	70.0	5.49	75.5	1.86	79.6	79.6	78.2	78.4	78.7	77.7	77.7	77.5	-	-	77.5	795	-
Vol. MSCFM at T°F	128.0	5.80	354.	1.97	370.	193.	216.	114.	154.	114.	137	153	-	-	230	840*	-
Vol. % CH_4	-	100.	-	100.	-	-	-	-	-	-	-	-	-	-	-	-	-
% SO_2	4.50	-	3.78	-	1.26	2.07	1.08	1.08	0.211	0.214	0.105	0.105	-	-	0.497	-	-
% H_2O	-	-	13.2	-	16.8	16.3	16.6	16.6	18.3	18.5	18.7	18.8	-	-	18.4		-
% H_2S	-	-	-	-	1.01	0.473	2.14	2.15	0.423	0.428	0.209	0.210	-	-	0.00	-	-
% CO_2	-	-	6.59	-	8.60	7.24	9.36	9.36	9.34	9.46	9.46	9.48	-	-	9.15	-	-
% CO	-	-	-	-	0.675	1.98	-	-	-	-	-	-	-	-	-	-	-
% S_2	-	-	-	-	0.847	0.697	0.414	0.197	1.48	0.199	0.363	0.100	-	-	0.00	-	100
% H_2	-	-	-	-	0.798	1.676	0.002	-	-	-	-	-	-	-	-	-	-
% COS	-	-	-	-	0.027	0.018	0.004	-	-	-	-	-	-	-	-	-	-
% N_2	79.0	-	66.4	-	70.0	69.5	70.4	70.6	70.3	70.2	71.2	71.4	-	-	72.0	-	-
% O_2	16.5	-	0.693	-	-	-	-	-	-	-	-	-	-	-	-	-	-

POINT	SULFUR DISTRIBUTION		
W	From Intermed. Converter Cond.	21.9	NT/Day
X	From No. 1 Claus Cond.	130.4	"
Y	From No. 2 Claus Cond.	26.4	"
S	Total (Incl. Coalescer)	178.7	"

* Includes CH_4 for H_2S incineration & preheaters

OVERALL CONVERSION 97.4 %

SULFUR YIELD 88.2 %

METHANE FOR RED'N: 15.0 MSCF/TON S

(15.8at 60°F)

For Fuel: 46.5 MCF/Ton S

METHANE REDUCTION OF SMELTER GAS

CASE 4c
GAS 2b

70,000 SCFM

KULCSAR PROCESS

Process Points	A	B	C	D	E	F	G	H	J	K	L	M	N	O	P	Q	S
Temp. °F	440	60	2093	60	2012	734	894	260	505	260	405	250	-	-	1200	60	250
Vol. MSCFM	70.0	5.49	75.5	1.86	79.2	79.6	72.2	78.4	78.7	77.7	77.7	77.5	-	-	77.5	795	-
Vol. MSCFM at T°F	128.0	5.80	391.	1.97	398.	193.	216.	114.	154.	114.	137.	153	-	-	230	840*	-
Vol. % CH_4	-	100.	-	100.	-	-	-	-	-	-	-	-	-	-	-	100	-
% SO_2	4.50	-	3.78	-	1.53	2.07	1.08	1.08	0.211	0.214	0.105	0.105	-	-	0.49	-	-
% H_2O	-	-	13.2	-	16.6	16.3	16.6	16.6	18.[illegible]	18.5	18.7	18.8	-	-	18.4	-	-
% H_2S	-	-	-	-	0.765	0.473	2.14	2.15	0.423	0.428	0.209	0.210	-	-	0.0[illegible]	-	-
% CO_2	-	-	6.59	-	8.12	7.24	9.36	9.36	9.34	9.46	9.46	9.48	-	-	9.15	-	-
% CO	-	-	-	-	1.14	1.98	-	-	-	-	-	-	-	-	-	-	-
% S_2	-	-	-	-	0.826	0.697	0.414	0.197	1.48	0.197	0.363	0.100	-	-	0.00	-	100.
% H_2	-	-	-	-	1.14	1.676	0.002	-	-	-	-	-	-	-	-	-	-
% COS	-	-	-	-	0.025	0.018	0.004	-	-	-	-	-	-	-	-	-	-
% N_2	79.0	-	66.4	-	69.8	69.5	70.4	70.6	70.3	70.2	71.2	71.4	-	-	72.0	-	-
% O_2	16.5	-	0.693	-	-	-	-	-	-	-	-	-	-		-		-

POINT	SULFUR DISTRIBUTION		
W	From Intermed. Converter Cond.	21.9	NT/Day
X	From No. 1 Claus Cond.	130.4	"
Y	From No. 2 Claus Cond.	26.4	"
S	Total (Incl. Coalescer)	178.7	"

* includes CH_4 for H_2S incineration & preheaters

OVERALL CONVERSION 97.4 %

SULFUR YIELD 88.0 %

METHANE FOR RED'N: 15.0 MSCF/TON S
(15.8 at 60°F)

For Fuel: 46.5 MCF/Ton S

Exhibit 10-7

SUMMARY - REACTOR DESIGN - CH_4 REDUCTION
BASIS: 5/16" dia. Porocel LPD (Tech. Info No. 1021)
Gas 2B

	High Temperature Reactors				Intermed.	Claus Unit	
Item	**West**	**Kulcsar(1)**	**Kulcsar(2)**	**Asarco**	**Converter**	**No. 1**	**No. 2**
Diameter, Ft.	33.8	21.6	33.8	24.0	33.8	45.0	45.0
Catalyst Depth, Ft.	10.0	1.9	0.77	10.0	10.5	6.95	6.88
Catalyst Volume, $Ft.^3$	9050.	697.	697.	4520.	9460.	11,050	10,930
SVH	500	6500	6500	1000	500	550	550
Superficial Veloc, Ft./Sec.,NTP	1.39	3.44	1.39	-	1.45	0.822	0.816
Δ P, PSI	1.40	1.40	0.115	0*	0.825	0.140	0.123
Δ P, Inches Water	38.8	38.8	3.19	0	22.8	3.88	3.41
Brake HP	1120	1120	92	0	658	112	98

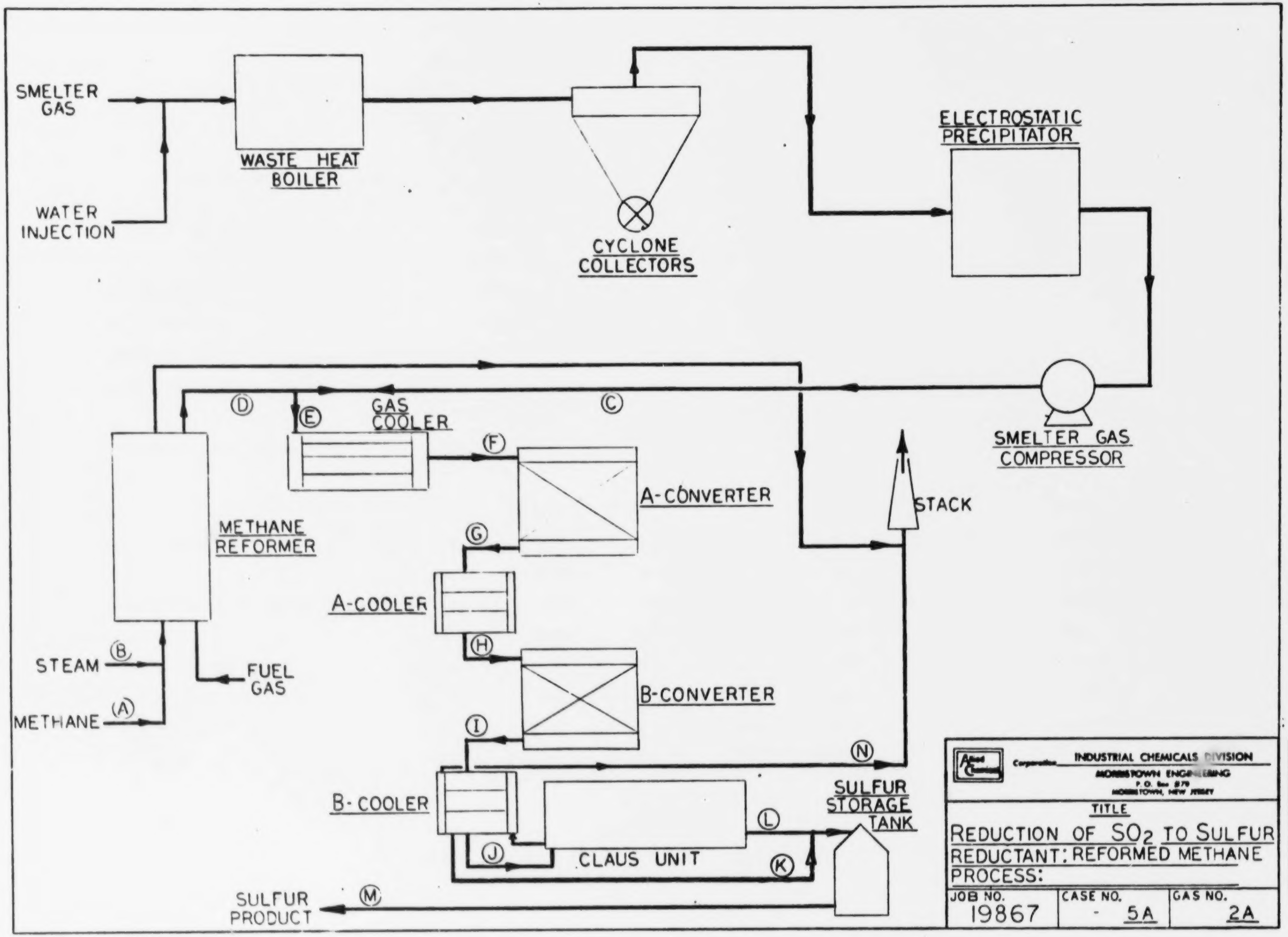
Compressor Duty 75,500 SCFM:

	West	**Kulcsar(1)**	**Kulcsar(2)**	**Asarco**
Primary Reactor	1.40	1.40	0.115	Neglig.
Intermed. Reactor	0.825	0.825	0.825	0.825
No. 1 Claus	0.140	0.140	0.140	0.140
No. 2 Claus	0.123	0.123	0.123	0.123
TotalΔ P, PSI	2.488	2.488	1.203	1.088
Brake HP, Total	1988	1988	960	858

*Kruzite Brick Checkerwork.

Exhibit 10-8

EXHIBIT 10-9


SMELTER GAS
WATER INJECTION
WASTE HEAT BOILER
CYCLONE COLLECTORS
ELECTROSTATIC PRECIPITATOR
SMELTER GAS COMPRESSOR
Ⓒ
Ⓓ
Ⓔ
GAS COOLER
Ⓕ
A-CONVERTER
STACK
METHANE REFORMER
Ⓖ
A-COOLER
Ⓗ
B-CONVERTER
STEAM
Ⓑ
METHANE
Ⓐ
FUEL GAS
Ⓘ
Ⓝ
B-COOLER
Ⓙ
CLAUS UNIT
Ⓚ
Ⓛ
SULFUR STORAGE TANK
Ⓜ
SULFUR PRODUCT

Allied Chemical Corporation INDUSTRIAL CHEMICALS DIVISION
MORRISTOWN ENGINEERING
P.O. Box 379
MORRISTOWN, NEW JERSEY

TITLE

REDUCTION OF SO_2 TO SULFUR
REDUCTANT: REFORMED METHANE
PROCESS:

JOB NO.	CASE NO.	GAS NO.
19867	- 5A	2A

DRAWING NO. 37196

EXHIBIT 10-10

REFORMED METHANE AND 2A SMELTER GAS

CASE 5-A

20000 SCFM SMELTER GAS

NO WATER REMOVAL FROM REFORMER GAS

PROCESS POINTS	A	B	C	D	E	F	G	H	I	J	N
TEMP. °C	25	205	260	898	560	300	650	340	572	205	150
VOL. SCFM	1735	5205	20000	10340	30340	30340	29230	29230	28120	27780	27030
VOL. CFM @ T°C	1895	9110	39040	44350	92570	63710	98830	65620	87030	48640	36760
% CH_4	100.00			0.35	0.12	0.12	0.12	0.12	0.13	0.13	0.13
% H_2O		100.00		29.03	9.89	9.89	17.11	17.11	24.89	25.19	31.29
% CO				11.57	3.94	3.94	2.05	2.05	0.00	0.00	0.00
% H_2				54.20	18.47	18.47	9.59	9.59	0.00	0.00	0.00
% CO_2				4.85	1.65	1.65	3.76	3.76	6.04	6.11	6.30
% SO_2			15.80		10.42	10.42	6.78	6.78	2.86	2.89	0.24
% O_2			1.20		0.79	0.79	0.41	0.41	0.00	0.00	0.00
% N_2			83.00		54.72	54.72	56.74	56.74	59.03	59.75	61.53
% H_2S							2.77	2.77	5.72	5.79	0.48
% S_2							0.64	0.64	1.33	0.00	0.00
% S_8										0.14	0.03

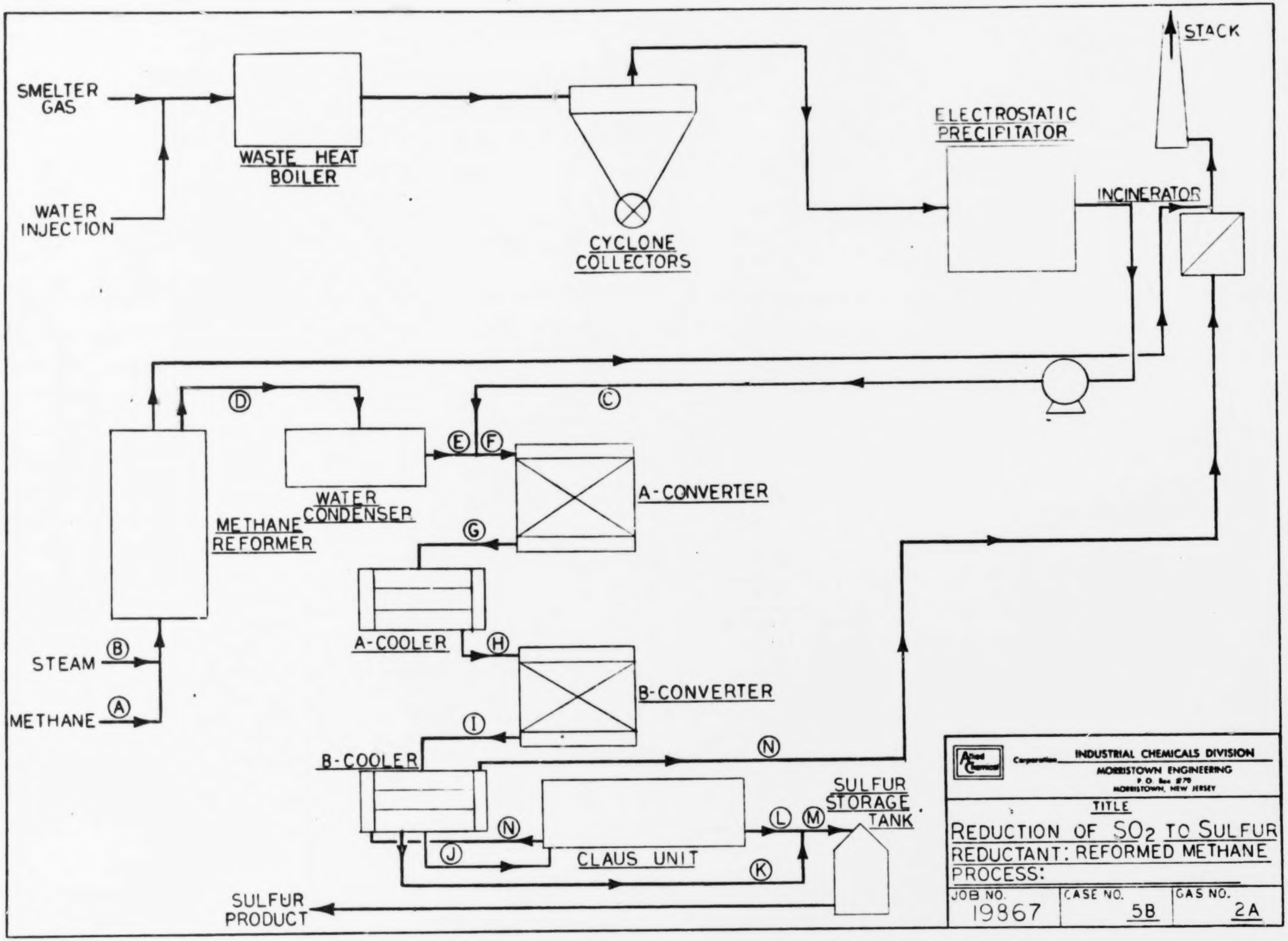
POINT	SULFUR DISTRIBUTION
Ⓚ	FROM "B" COOLER 28.98 NT/DAY
Ⓛ	FROM CLAUS 155.55 " "
Ⓜ	TOTAL SULFUR 184.53 " "

OVERALL CONVERSION 93.91 %

SULFUR YIELD 92.00 %

METHANE FACTOR 13540 SCF/TON SULFUR

EXHIBIT 10-11


SMELTER GAS
WATER INJECTION
WASTE HEAT BOILER
CYCLONE COLLECTORS
ELECTROSTATIC PRECIPITATOR
INCINERATOR
STACK
STEAM
Ⓑ
METHANE
Ⓐ
METHANE REFORMER
Ⓓ
WATER CONDENSER
Ⓔ
Ⓕ
Ⓒ
A-CONVERTER
Ⓖ
A-COOLER
Ⓗ
B-CONVERTER
Ⓘ
B-COOLER
Ⓝ
Ⓙ
CLAUS UNIT
Ⓚ
Ⓛ
Ⓜ
SULFUR STORAGE TANK
SULFUR PRODUCT

Allied Chemical Corporation INDUSTRIAL CHEMICALS DIVISION
MORRISTOWN ENGINEERING
P.O. Box 879
MORRISTOWN, NEW JERSEY

TITLE
REDUCTION OF SO_2 TO SULFUR
REDUCTANT: REFORMED METHANE
PROCESS:

JOB NO.	CASE NO.	GAS NO.
19867	5B	2A

DRAWING NO. 37197

REFORMED METHANE AND CA SMELTER GAS

CASE 5B

20000 SCFM SMELTER GAS
WATER REMOVAL FROM REFORMER GAS

PROCESS POINTS	A	B	C	D	E	F	G	H	I	J	N
TEMP. °C	25	205	390	898	60	300	700	355	600	205	150
VOL. SCFM	1735	5205	20000	10340	7490	27490	26420	26420	25350	24760	24120
VOL. CFM @ T°C	1895	9110	48580	44350	9140	57730	94500	60770	81070	43350	32800
% CH_4	100.00			0.35	0.48	0.13	0.14	0.14	0.14	0.14	0.15
% H_2O		100.00		29.03	1.99	0.54	8.78	8.78	17.71	18.13	23.43
% CO				11.57	15.98	4.35	2.26	2.26	0.00	0.00	0.00
% H_2				54.20	74.85	20.39	10.61	10.61	0.00	0.00	0.00
% CO_2				4.85	5.70	1.83	4.16	4.16	6.70	6.86	7.04
% SO_2			15.80			11.50	7.18	7.18	2.49	2.55	0.18
% O_2			1.20			0.87	0.46	0.46	0.00	0.00	0.00
% H_2S							2.39	2.39	4.98	5.11	0.36
% N_2			83.00			60.39	62.83	62.83	65.49	67.06	68.81
% S_2							1.19	1.19	2.49		
% S_8										0.15	0.03

POINT	SULFUR DISTRIBUTION	
Ⓚ	FROM "B" COOLER	61.70 NT/DAY
Ⓛ	FROM CLAUS	129.06 " "
Ⓜ	TOTAL SULFUR	190.76 " "

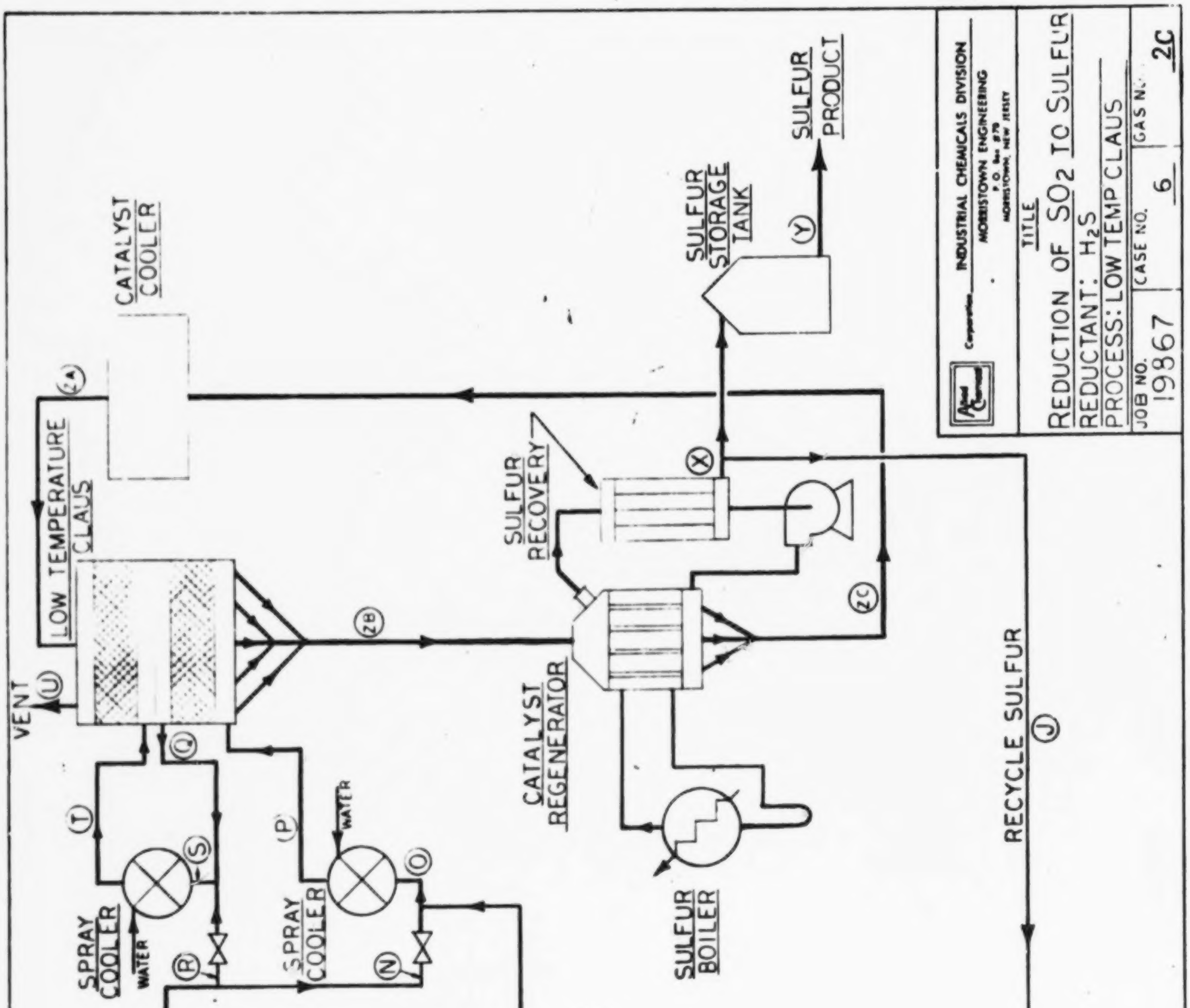
OVERALL CONVERSION 95.88 %

SULFUR YIELD 93.97 %

METHANE FACTOR 13100 SCF / TON SULFUR

EXHIBIT 10-12

EXHIBIT 10-13


SMELTER GAS
Ⓜ
SMELTER GAS COMPRESSOR
Ⓚ
H_2S COOLER
INTER COOLER
Ⓕ
Ⓔ
Ⓓ
Ⓒ
METHANE REFORMER
METHANE
Ⓐ
STEAM
Ⓑ
Ⓘ
PUMP TANK
SULFUR COOLER
SULFUR PUMP
SPRAY COOLER
WATER
Ⓡ
Ⓢ
Ⓣ
SPRAY COOLER
WATER
Ⓝ
Ⓞ
Ⓟ
Ⓠ
VENT
Ⓤ
LOW TEMPERATURE CLAUS
(ZA)
CATALYST COOLER
(ZB)
CATALYST REGENERATOR
SULFUR BOILER
(ZC)
SULFUR RECOVERY
Ⓧ
RECYCLE SULFUR
Ⓙ
SULFUR STORAGE TANK
Ⓨ
SULFUR PRODUCT

Corporation
INDUSTRIAL CHEMICALS DIVISION
MORRISTOWN ENGINEERING
P. O. Box 879
MORRISTOWN, NEW JERSEY

TITLE
REDUCTION OF SO_2 TO SULFUR
REDUCTANT: H_2S
PROCESS: LOW TEMP CLAUS

JOB NO.	CASE NO.	GAS NO.
19867	6	2C

DRAWING NO. 37198

EXHIBIT 10-14

LOW TEMPERATURE CLAUS

210000 SCFM 2C SMELTER GAS

EXH. 10-14

CASE 6

PROCESS POINTS	A	B	C	D	E	F	K	L		M	N	O	P	Q	R	S	T	U	V	W
TEMP. °C	25	205	898	300	625	300	540	125		227	125	223	65	115	125	116	65	115	125	420
SCFM	3073	9217	18320	19850	19300	19300	18640	18580		210000	9290	219290	243290	234200	9290	243490	252540	249500	1000	3160
CFM @T°C	3360	16140	78570	41690	63480	40530	55510	27090		384720	13540	398450	301190	332800	13540	346980	312640	354540	1460	8020
DEW POINT °C													47.5				53.4			
% CH_4	100.00		0.35	0.32	0.11	0.11														
% H_2O		100.00	29.03	26.78	19.03	19.03	16.74	16.80		0.10	16.80	0.81	10.60	11.01	16.80	11.23	14.40	16.98		
% CO			11.57	10.68	3.96	3.96	0.11	0.11		0.60	0.11	0.58	0.52	0.54	0.11	0.53	0.51	0.51		
% H_2			54.20	50.00	19.05	19.05	1.31	1.31			1.31	0.05	0.05	0.05	1.31	0.10	0.10	0.10		
% CO_2			4.85	4.48	10.53	10.53	16.17	16.23		1.70	16.23	2.32	2.09	2.17	16.23	2.70	2.61	2.67		
% COS					0.15	0.15	0.34	0.34			0.34	0.02	0.02	0.00	0.34	0.02	0.02	0.00		
% H_2S					35.69	35.69	64.98	65.21			65.21	2.76	2.50	0.00	65.21	2.50	2.40	0.024		
% S_2							0.34	0.00			0.00									S_6 37.70
% S_8				7.74																30.80
% SO_2										2.90		2.78	2.50	1.31		1.26	1.21	0.012		
% O_2										14.30		13.68	12.34	12.82		12.33	11.89	12.03		
% N_2										80.40		77.00	69.38	72.08		69.34	66.86	67.67	100.00	31.50

POINT	SULFUR DISTRIBUTION	MT./DAY
Ⓧ	FROM CATALYST	1166.72
Ⓘ	FROM GAS COOLER	8.15
	TOTAL S RECOVERY	1174.87
Ⓙ	RECYCLE SULFUR	789.83
Ⓨ	NET S RECOVERY	385.04

CONVERSION

EQULIBRIUM 99.63%

USED HERE 99.50%

SULFUR YIELD 98.52%

METHANE FACTOR 11490 SCF/TON S

CATALYST CIRCULATION RATE 194.4 TON/Hr.

POINT	TEMP °C
G	300
H	250
ZA	100
ZB	75
ZC	400

EXHIBIT 10-15
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: CH_4 CASE NO. 3a
PROCESS: Asarco GAS NO. 2A

Fixed Capital, Battery Limits Basis $ 3,260,000
Sulfur Production 194 NTD x 330 Days = 64,000 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	13.4 MCF	$.30 Avg.	$ 4.02	$ 257.0
1.	Total Raw Materials	-	-	$ 4.02	$ 257.0
2.	Direct Labor	18,000 MH	$3.50	.98	63.0
3.	Supervision	-	$10,000	.16	10.0
4.	Maintenance	5% FC	-	2.54	163.0
5.	Plant Supplies	10% (L2+L3)	-	.11	7.3
	Utilities:				
6. a)	Electricity	181 KWH	$.01	1.81	116.0
b)	Cooling Water	3.62 MG	.05	.18	11.5
c)	Boiler Feed Water	0.90 MG	.25	.23	14.4
d)	Fuel Gas	5.0 MCF	.30	1.50	96.0
7.	Total Utilities	-	-	3.72	237.9
8.	Other Direct (Catalyst)	0.8 Lbs.	.13	.10	6.4
9.	TOTAL DIRECT	-	-	$ 11.63	$ 744.6
10.	Payroll Burden	30% L2	-	.30	18.9
11.	Plant Overhead	70% (L2+L3)	-	.80	51.1
12.	Pack & Ship	1000 MH	3.90	.06	3.9
13.	Waste Disposal	1000 MH	3.90	.06	3.9
14.	Other Indirect	-	-	-	-
15.	TOTAL INDIRECT	-	-	$ 1.22	$ 77.8
16.	Depreciation	10% FC	-	5.10	326.0
17.	Taxes	2% FC	-	1.02	65.2
18.	Insurance	1% FC	-	.51	32.6
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$ 6.63	$ 423.8
21.	TOTAL OPERATING	-	-	$ 19.48	$ 1,246.2
22.	Credits HP Steam	3.46 M Lb.	.50	1.73	111.0
23.	NET OPERATING	-	-	$ 17.75	$ 1,135.2

EXHIBIT 10-16

METHANE REDUCTION OF SO_2
2A SMELTER GAS
CASE 3A ASARCO PROCESS
SUMMATION OF MAJOR EQUIPMENT

1.	Hot Gas Ductwork - 2300°F smelter gas, 20000 SCRM, is reduced to 2000°F by direct water spray injection at the converter outlets and passed through refractory lined steel ducts to waste heat boiler. Allowance for cost of this ductwork, as a major equipment item.	$ 50,000
2.	Waste Heat Boiler - 40,000 lb./hr. high pressure steam, water tube design, with soot blowers and auxiliaries.	$ 80,000
3.	Cyclone Collector System - multiple steel cyclone collectors with ash gates.	$ 10,000
4.	Precipitator - rod curtain design electrostatic precipitator, carbon steel construction, for 21000 SCFM 500°F gas.	$ 40,000
5.	Smelter Gas Blower - centrifugal blower to handle 21000 SCFM at an inlet temperature of 500°F with a ΔP of 2 psi. Complete with a 500 HP turbine driver.	$ 45,000
6.	Steam Condenser - surface condenser for exhaust steam from smelter gas compressor turbine.	$ 5,000
7.	Reactor-Regenerator - carbon steel refractory brick lined chamber segmented into two regenerator sections and one reactor section by means of refractory division walls. Each regenerator consists of 2000 cubic feet of packed refractory material on a refractory grillwork. Reactor section consists of an 8 ft. deep Kruzite refractory checkerwork with approximately 50% void space, supported on a similar refractory grillwork.	$180,000
8.	Flow reversing valve - refractory lined four-way piston valve	$ 50,000
9.	Heat Exchanger - shell and tube process water cooler of stainless steel construction.	$ 30,000
10.	Intermediate Converter - stainless steel conversion chamber approximately 20 ft. diameter by 20 ft. high complete with 3000 cu. ft. of alumina catalyst supported on a cast iron grillwork, externally insulated.	$ 90,000
11.	Primary Condenser - low pressure steam generator designed to condense sulfur from 25000 SCFM of 730°F gas at a rate of 136 short tons per day.	$ 70,000

EXH. 10-16 (Cont.)

12.	Process Blower - centrifugal pressure blower for 24,000 SCRM at ΔP of 5 psig equipped with a 1500 HP motor.	$110,000
13.	Gas Preheaters (2) - Natural gas fired auxiliary burners, each designed to reheat Claus process gas stream for feed to Claus converters. Preheater outlet temperature 450°F. Cost for two units.	$ 20,000
14.	Claus Converters - carbon steel refractory lined vessel 14 ft. diameter by 16 ft. high with two converter sections, each having 300 cubic ft. alumina catalyst. Complete with grillwork for catalyst support.	$ 30,000
15.	Sulfur Condensers (2) - shell and tube carbon steel condensers to recover a total of 58 short tons of sulfur per day from process gas stream. Units preheat boiler feedwater for use in waste heat boiler item 2 above.	$100,000
16.	Coalescer - carbon steel circulating liquid sulfur type coalescer of standard design, complete with auxiliaries including pump.	$ 20,000
17.	Incinerator and Stack - carbon steel refractory brick lined incinerator and 150 foot self supporting refractory lined stack, to heat process effluent to 1200°F and discharge.	$120,000
18.	Sulfur Storage Tank - vertical 2000 ton capacity with steam coils and external insulation.	$ 40,000
Total Purchased Equipment		$1,090,000
Plant Cost		$3,260,000

EXHIBIT 10-17
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: CH_4 CASE NO. 3b
PROCESS: West (TGS) GAS NO. 2A

Fixed Capital, Battery Limits Basis $ 3,500,000
Sulfur Production 194 NTD x 330 Days = 64,000 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	13.4 MCF	$.30 Avg.	$ 4.02	$ 257.0
1.	Total Raw Materials	-	-	$ 4.02	$ 257.0
2.	Direct Labor	18,000MH	$3.50	.98	63.0
3.	Supervision	1.	$10,000	.16	10.0
4.	Maintenance	5% FC	-	2.74	175.0
5.	Plant Supplies	10%(L2+L3)	-	.11	7.3
	Utilities:				
6. a)	Electricity	181 KWH	$.01	1.81	116.0
b)	Cooling Water	3.62 MG	.05	.18	11.5
c)	Boiler Feed Water	0.86 MG	.25	.22	13.8
d)	Fuel Gas	5.0 MCF	.30	1.50	96.0
7.	Total Utilities	-	-	3.71	237.3
8.	Other Direct (Catalyst)	3.8 Lbs.	.13	.49	31.6
9.	TOTAL DIRECT	-	-	$ 12.21	$ 781.2
10.	Payroll Burden	30% L2	-	.30	18.9
11.	Plant Overhead	70%(L2+L3)	-	.80	51.1
12.	Pack & Ship	1000 MH	3.90	.06	3.9
13.	Waste Disposal	1000 MH	3.90	.06	3.9
14.	Other Indirect	-	-	-	-
15.	TOTAL INDIRECT	-	-	$ 1.22	$ 77.8
16.	Depreciation	10% FC	-	5.48	350.0
17.	Taxes	2% FC	-	1.10	70.0
18.	Insurance	1% FC	-	.55	35.0
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$ 7.13	$ 455.0
21.	TOTAL OPERATING	-	-	$ 20.56	$ 1,314.0
22.	Credits HP Steam	3.2 M Lb.	.50	1.60	102.0
23.	NET OPERATING	-	-	$ 18.96	$ 1,212.0

EXHIBIT 10-18

METHANE REDUCTION OF SO_2
2A SMELTER GAS
CASE 3B WEST PROCESS
SUMMATION OF MAJOR EQUIPMENT

1.	Hot Gas Ductwork - 2300°F smelter gas, 20000 SCFM, is reduced to 2000°F by direct water spray injection at the converter outlets and passed through refractory lined steel ducts to waste heat boiler. Allowance for cost of this ductwork, as a major equipment item.	$ 50,000
2.	Waste Heat Boiler - 40,000 lb./hr. high pressure steam, water tube design, with soot blowers and auxiliaries.	$ 80,000
3.	Cyclone Collector System - multiple steel cyclone collectors with ash gates.	$ 10,000
4.	Precipitator - rod curtain design electrostatic precipitator, carbon steel construction, for 21000 SCFM 500°F gas.	$ 40,000
5.	Smelter Gas Blower - centrifugal blower to handle 21000 SCFM at an inlet temperature of 500°F with a ΔP of 2 1/2 psi. Complete with a 600 HP turbine driver.	$ 50,000
6.	Steam Condenser - surface condenser for exhaust steam from smelter gas compressor turbine.	$ 5,000
7.	Reactor-Regenerator - carbon steel refractory brick lined chamber segmented into two regenerator sections and one reactor section, connected by external ductwork and hot gas valves. Each regenerator consists of 2000 cubic ft. of packed refractory material on a refractory grillwork. The reactor section consists of 3100 cubic feet alumina type catalyst supported on a refractory grillwork. Cost including hot gas valves.	$250,000
8.	Flow Reversing Valve - refractory lined four-way piston valve.	$ 50,000
9.	Heat Exchanger - shell and tube process water cooler of stainless steel construction.	$ 30,000
10.	Intermediate Converter - stainless steel conversion chamber approximately 20 ft. diameter by 20 ft. high complete with 3000 cu. ft. of alumina catalyst supported on a cast iron grillwork, externally insulated.	$ 90,000
11.	Primary Condenser - low pressure steam generator designed to condense sulfur from 25000 SCFM of 730°F gas at a rate of 132 short tons per day.	$ 70,000

EXH. 10-18 (Cont.)

12.	Process Blower - centrifugal pressure blower for 24000 SCFM at ΔP of 5 psig equipped with a 1500 HP motor.	$110,000
13.	Gas Preheaters (2) - Natural gas fired auxiliary burners, each designed to reheat Claus process gas stream for feed to Claus converters. Preheater outlet temperature 450°F. Cost for two units.	$ 20,000
14.	Claus Converters - carbon steel refractory lined vessel 14 ft. diameter by 16 ft. high with two converter sections, each having 300 cubic ft. alumina catalyst. Complete with grillwork for catalyst support.	$ 30,000
15.	Sulfur Condensers (2) - shell and tube carbon steel condensers to recover a total of 60 short tons of sulfur per day from process gas stream. Units preheat boiler feedwater for use in waste heat boiler item 2 above.	$100,000
16.	Coalescer - carbon steel circulating liquid sulfur type coalescer of standard design, complete with auxiliaries including pump.	$ 20,000
17.	Incinerator and Stack - carbon steel refractory brick lined incinerator and 150 foot self supporting refractory lined stack, to heat process effluent to 1200°F and discharge.	$120,000
18.	Sulfur Storage Tank - vertical 2000 ton capacity with steam coils and external insulation.	$ 40,000
Total purchased equipment.		$1,165,000
Plant Cost		$3,500,000

EXHIBIT 10-19
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: Reformed CH_4 CASE NO. 5a
PROCESS: Wet Gas GAS NO. 2A
Fixed Capital, Battery Limits Basis $ 3,000,000
Sulfur Production 185 NTD x 330 Days = 61,000 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	143 MCF	$.30 Avg.	$ 4.30	$ 262.0
1.	Total Raw Materials	-	-	$ 4.30	$ 262.0
2.	Direct Labor	18,000 MH	$3.50	1.03	63.0
3.	Supervision	-	$10,000	.16	10.0
4.	Maintenance	5% FC	-	2.46	150.0
5.	Plant Supplies	10%(L2+L3)	-	.12	7.3
	Utilities:				
6. a)	Electricity	65 KWH	$.01	.65	39.6
b)	Cooling Water	6.10 MG	.05	.30	18.3
c)	Boiler Feed Water	1.06 MG	.25	.27	16.5
d)	Fuel Gas	10.4 MCF	.30	3.12	190.0
7.	Total Utilities	-	-	4.34	264.4
8.	Other Direct (Catalyst)	3.6 Lbs.	.13	.47	28.6
9.	TOTAL DIRECT	-	-	$ 12.88	$ 785.3
10.	Payroll Burden	30% L2	-	.31	18.9
11.	Plant Overhead	70%(L2+L3)	-	.84	51.1
12.	Pack & Ship	1000 MH	3.90	.06	3.9
13.	Waste Disposal	1000 MH	3.90	.06	3.9
14.	Other Indirect	-	-	-	-
15.	TOTAL INDIRECT	-	-	$ 1.27	$ 77.8
16.	Depreciation	10% FC	-	4.92	300.0
17.	Taxes	2% FC	-	.98	60.0
18.	Insurance	1% FC	-	.49	30.0
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$ 6.39	$ 390.0
21.	TOTAL OPERATING	-	-	$ 20.54	$ 1,253.1
22.	Credits HP Steam	6.25 M Lb.	.50	3.12	190.0
23.	NET OPERATING	-	-	$ 17.42	$ 1,063.1

EXHIBIT 10-20

REFORMED METHANE REDUCTION OF SO_2
2A SMELTER GAS
CASE 5A
SUMMATION OF MAJOR EQUIPMENT

1.	Hot gas ductwork - 2300°F smelter gas, 20000 SCFM, is reduced to 2000°F by direct water spray injection at the converter outlets and passed through refractory lined steel ducts to waste heat boiler. Allowance for cost of this ductwork, as a major equipment item.	$ 50,000
2.	Waste heat boiler - 40,000 lb./hr. high pressure steam, water tube design, with soot blowers and auxiliaries.	$ 80,000
3.	Cyclone collector system - multiple steel cyclone collectors with ash gates.	$ 10,000
4.	Precipitator - rod curtain design electrostatic precipitator, carbon steel construction, for 21000 SCFM 500°F gas.	$ 40,000
5.	Smelter Gas Compressor - two stage centrifugal compressor to handle 21,000 SCFM at an inlet temperature of 500°F with a ΔP of 7 psi. Complete with a 1200 HP turbine driver.	$110,000
6.	Steam condenser - surface condenser for exhaust steam from smelter gas compressor turbine.	$ 10,000
7.	Reformer - direct gas fired box type reforming furnace designed to reform 1750 SCFM methane at 1600°F, 150 psig outlet incorporating charge preheat coils and high pressure steam generation coils in convection section. Thermal efficiency 85% including economizer coils. High pressure steam generated is used to operate turbine for item 5 above. Complete with twin carbon bed desulfurizers.	$260,000
8.	Gas Cooler - two stage gas to gas alonized steel shell and tube heat exchanger used to cool smelter gas and reformer gas blend. First stage exchanges against reformer stack gas. Second stage trim cools using cold air indirect cooling. Complete with air blower.	$ 40,000
9.	Converters "A" and "B" - carbon steel 15 ft. diameter by 25 ft. high vertical vessel of two compartments each having 6 ft. catalyst bed, support gratings and refractory lining.	$ 60,000
10.	Cooler "A" - interstage cooler for converters, medium pressure steam generator 21 MM BTU/hr. producing 250 psig steam which is used as charge to reformer item 7 above, fire tube design.	$ 40,000

11.	Cooler "B" - precooler for Claus unit reducing process gas stream to 400°F. Shell and tube gas to gas design exchanging against Claus unit tail gas, preheating this tail gas for release to atmosphere. Stainless steel tubes, alonized shell.	$ 80,000
12.	Claus Unit - partial Claus sulfur recovery unit consisting of two stage converter, two sulfur condensers, coalescer and exit stack. Unit is designed for a throughput of 28000 SCFM. Sulfur condensers preheat feedwater for waste heat boilers items 2, and 10 above, and for economizer section of reformer item 7. Tail gas from coalescer is heated in exchanger item 11, then mixed with reheated reformer off gas from item 8. Mixed high temperature off gases are released through stack. No incineration is provided.	$300,000
13.	Sulfur storage tank - vertical steel tank 2000 ton capacity with steam coils and external insulation.	$ 40,000
	Total Purchased Equipment	$1,120,000
	Plant Cost	$3,000,000

EXHIBIT 10-21
PH 22-68-24

REDUCTION OF SO$_2$

REDUCTANT: Reformed CH$_4$ CASE NO. 5b
PROCESS: Dry Gas GAS NO. 2A

Fixed Capital, Battery Limits Basis $ 3,000,000
Sulfur Production 190 NTD x 330 Days = 62,700 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	13.8 MCF	$.30 Avg.	$ 4.13	$ 259.0
1.	Total Raw Materials	-	-	$ 4.13	$ 259.0
2.	Direct Labor	18,000 MH	$3.50	1.00	63.0
3.	Supervision		$10,000	.16	10.0
4.	Maintenance	5% FC	-	2.39	150.0
5.	Plant Supplies	10%(L2+L3)	-	.12	7.3
	Utilities:				
6. a)	Electricity	63 KWH	$.01	.63	39.5
b)	Cooling Water	8.60 MG	.05	.43	27.0
c)	Boiler Feed Water	1.29 MG	.25	.32	20.1
d)	Fuel Gas	10.6 MCF	.30	3.18	199.0
7.	Total Utilities	-	-	4.56	285.6
8.	Other Direct (Catalyst)	3.6 Lbs.	.13	.47	29.4
9.	TOTAL DIRECT	-	-	$ 12.83	$ 804.3
10.	Payroll Burden	30% L2	-	.30	18.9
11.	Plant Overhead	70%(L2+L3)	-	.82	51.1
12.	Pack & Ship	1000 MH	3.90	.06	3.9
13.	Waste Disposal	1000 MH	3.90	.06	3.9
14.	Other Indirect	-	-	-	-
15.	TOTAL INDIRECT	-	-	$ 1.24	$ 77.8
16.	Depreciation	10% FC	-	4.78	300.0
17.	Taxes	2% FC	-	.96	60.0
18.	Insurance	1% FC	-	.48	30.0
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$ 6.22	$ 390.0
21.	TOTAL OPERATING	-	-	$ 20.29	$ 1,272.1
22.	Credits HP Steam	8.2 M Lb.	.50	4.10	257.0
23.	NET OPERATING	-	-	$ 16.19	$ 1,015.1

EXHIBIT 10-22

REFORMED METHANE REDUCTION OF SO_2
2A SMELTER GAS
CASE 5b
SUMMATION OF MAJOR EQUIPMENT

1. Hot gas ductwork - 2300°F smelter gas, 20000 SCFM, is reduced to 2000°F by direct water spray injection at the converter outlets and passed through refractory lined steel ducts to waste heat boiler. Allowance for cost of this ductwork, as a major equipment item.---------------- $ 50,000

2. Waste heat boiler - 33,000 lb./hr. high pressure steam, water tube design, with soot blowers and auxiliaries.----- $ 70,000

3. Cyclone collector system - multiple steel cyclone collectors with ash gates.-------------------------------- $ 10,000

4. Precipitator - rod curtain design electrostatic precipitator, carbon steel construction, for 21000 SCFM 500°F gas.------ $ 40,000

5. Smelter Gas Compressor - two stage centrifugal compressor to handle 21000 SCFM at an inlet temperature of 735°F with a ΔP of 7 psi. Complete with a 1200 HP turbine driver.--- $110,000

6. Steam condenser - surface condenser for exhaust steam from smelter gas compressor turbine.------------------------- $ 10,000

7. Reformer - direct gas fired box type reforming furnace designed to reform 1750 SCFM methane at 1600°F, 150 psig outlet incorporating charge preheat coils and high pressure steam generation coils in connection section. Thermal efficiency 85% including economizer coils. High pressure steam generated is used to operate turbine for item 5, above. Complete with twin carbon bed desulfurizers.-------------- $260,000

8. Gas Cooler - two stage cooler consisting of fire tube boiler to reduce gas temperature to 500°F, generating 22,000 lb./hr. high pressure steam, followed by water quench chamber to reduce gas temperature to 140°F.------------------------ $ 40,000

9. Converters "A" and "B" - carbon steel 14 ft. diameter by 25 ft. high vertical vessel of two compartments each having 6 ft. catalyst bed, support gratings and refractory lining.--- $ 50,000

10. Cooler "A" - interstage cooler for converters, medium pressure steam generator 22 MM BTU/hr. producing 250 psig steam which is used as charge to reformer item 7 above, fire tube design.-- $ 40,000

EXH. 10-22 (Cont.)

11.	Cooler "B" - precooler for Claus unit reducing process gas stream to 400°F. Shell and tube gas to gas design exchanging against Claus unit tail gas, preheating this tail gas for release to atmosphere. Stainless steel tubes, alonized shell.	$ 80,000
12.	Claus unit - partial Claus sulfur recovery unit consisting of two stage converter, two sulfur condensers, coalescer, incinerator and exit stack. Unit is designed for a throughput of 26,500 SCFM. Sulfur condensers preheat feedwater for waste heat boilers items 2 and 10 above, and for economizer section of reformer item 7. Tail gas from coalescer is heated in exchanger item 11, then further incinerated to 1200°F before mixing with reformer off gas for exhaust to atmosphere.	$320,000
13.	Sulfur storage tank - vertical steel tank 2000 ton capacity with steam coils and external insulation.	$ 40,000
Total Purchased Equipment		$1,120,000
Plant Cost		$3,000,000

EXHIBIT 10-23
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: CH_4 CASE NO. 4a
PROCESS: Asarco GAS NO. 2B

Fixed Capital, Battery Limits Basis $ 9,150,000
Sulfur Production 180 NTD x 330 Days = 59,000 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	62.2 MCF	$.30 Avg.	$ 18.66	$ 1,100.0
1.	Total Raw Materials	-	-	$ 18.66	$ 1,100.0
2.	Direct Labor	18,000 MH	$3.50	1.07	63.0
3.	Supervision	1.	$10,000	.17	10.0
4.	Maintenance	5% FC	-	7.75	457.5
5.	Plant Supplies	10%(L2+L3)	-	.12	7.3
	Utilities:				
6. a)	Electricity	70 KWH	$.01	.70	41.3
b)	Cooling Water	41.8 MG	.05	2.09	123.2
c)	Boiler Feed Water	2.42 MG	.25	.61	36.0
d)	Fuel Gas	7.10 MCF	.30	2.13	125.8
7.	Total Utilities	-	-	5.53	326.3
8.	Other Direct (Catalyst)	9 Lbs.	.13	1.17	69.0
9.	TOTAL DIRECT	-	-	$ 33.30	$ 1,964.1
10.	Payroll Burden	30% L2	-	.32	18.9
11.	Plant Overhead	70%(L2+L3)	-	.87	51.1
12.	Pack & Ship	1000 MH	3.90	.07	3.9
13.	Waste Disposal	1000 MH	3.90	.07	3.9
14.	Other Indirect	-	-	-	-
15.	TOTAL INDIRECT	-	-	$ 1.33	$ 77.8
16.	Depreciation	10% FC	-	15.50	915.0
17.	Taxes	2% FC	-	3.10	183.0
18.	Insurance	1% FC	-	1.55	91.5
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$ 20.15	$ 1,189.5
21.	TOTAL OPERATING	-	-	$ 54.78	$ 3,231.4
22.	Credits HP Steam	15.9 M Lb.	.50	7.95	469.0
23.	NET OPERATING	-	-	$ 46.83	$ 2,762.4

EXHIBIT 10-24

METHANE REDUCTION OF SO_2
CASE 4a
ASARCO PROCESS
GAS 2B
SUMMATION OF MAJOR EQUIPMENT

1.	Cyclone collector system - multiple steel cyclone collectors with ash gates.	$ 40,000
2.	Precipitator - rod curtain design electrostatic precipitator, carbon steel construction for 70,000 SCFM 440°F gas.	$150,000
3.	Primary Blower - centrifugal gas blower to handle 70,000 SCFM gas at 440°F with a Δ P of 3 psi complete with 3,000 HP condensing steam turbine driver.	$250,000
4.	Waste Heat Boiler #1 - forced circulation water tube boiler with superheater with combustion chamber for methane - smelter gas reaction generating 66,000 lbs./hr. high pressure steam.	$180,000
5.	Reactor - Brick-lined carbon steel reaction vessel 24 ft. diameter, 30 ft. high with 10 ft. Kruzite brick checkerwork catalyst section supported on refractory grillwork	$ 80,000
6.	Waste Heat Boiler #2 - water tube boiler with superheater generating 176,000 lb./hr. high pressure superheated steam.	$400,000
7.	Intermediate Reactor - stainless steel conversion chamber 34 ft. diameter by 25 ft. high complete with 9,500 cu. ft. alumina catalyst supported on a cast iron grillwork, externally insulated.	$260,000
8.	Primary Condenser - low pressure steam generator of carbon steel construction designed to condense sulfur from 78,500 SCFM gas at 894°F at a rate of 22 short tons per day. Low pressure steam generation 67,000 lbs./hr.	$210,000
9.	Gas Preheater #1 Shell and tube steam heater for 78,000 SCFM raising gas to 450°F for Claus #1 inlet. Heat duty 20 MM BTU/hr. using high pressure steam.	$ 90,000
10.	Claus Unit Two stage Claus converter consisting of 45 ft. diameter carbon steel vertical vessel with two 7 ft. deep catalyst beds. Overall height 30 ft. Complete with 22,000 cubic feet alumina type catalyst and supports.	$370,000

EXH. 10-24 (Cont.)

11.	Process blower - Two stage centrifugal pressure blower for 4 PSIG P 78,000 SCFM at 260°F, complete with 4,000 HP turbine driver.----------------------------------	$280,000
12.	Steam condensers - condensers for steam from turbine drivers for items 3 and 11 above, complete with auxiliaries.-----	40,000
13.	Gas Preheater #2 Shell and tube steam heater for 78,000 SCFM raising gas to 400°F for Claus #2 inlet. Heat duty 14 MM BTU/hr. using high pressure steam.-------------------------------	$ 70,000
14.	Claus condensers (2) Shell and tube exchangers condensing sulfur from outlets of two stages of Claus converters. Boiler feedwater is preheated in the exchanger. Cost for two units.---------	$220,000
15.	Coalescer - carbon steel circulating liquid sulfur type coalescer of standard design, complete with auxiliaries including pump.--	$ 40,000
16.	Tail Gas Heater - shell and tube steam heater for tail gas prior to incineration, raising temperature to 700°F by exchange with superheated steam.------------------------	$100,000
17.	Incinerator and stack - carbon steel refractory-lined stack with lower section fitted as auxiliary heater to raise temperature of tail gas to 1200°F and discharge, complete with auxiliaries.------------------------------	$200,000
18.	Sulfur Storage Tank - vertical 5,000 ton capacity with steam coils and external insulation.--------------------	$ 70,000

Total Purchased Equipment.-----------------------------------$2,040,000

Total Installed Fixed Capital.--------------------------------$9,140,000

EXHIBIT 10-25
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: CH_4 CASE NO. 4b
PROCESS: West (TGS) GAS NO. 2B

Fixed Capital, Battery Limits Basis $9,800,000
Sulfur Production 180 NTD x 330 Days = 59,000 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	62.2 MCF	$.30 Avg.	$ 18.66	$ 1,100.0
1.	Total Raw Materials	-	-	$ 18.66	$ 1,100.0
2.	Direct Labor	18,000 MH	$3.50	1.07	63.0
3.	Supervision	1.	$10,000	.17	10.0
4.	Maintenance	5% FC	-	8.31	490.0
5.	Plant Supplies	10%(L2+L3)	-	.12	7.3
	Utilities:				
6. a)	Electricity	70 KWH	$.01	.70	41.3
b)	Cooling Water	45.0 MG	.05	2.25	132.5
c)	Boiler Feed Water	0.98 MG	.25	.25	14.8
d)	Fuel Gas	7.10 MCF	.30	2.13	125.6
7.	Total Utilities	-	-	5.33	314.2
8.	Other Direct (Catalyst)	16.6 Lbs.	.13	2.16	127.0
9.	TOTAL DIRECT	-	-	$ 35.82	$ 2,111.5
10.	Payroll Burden	30% L2	-	.32	18.9
11.	Plant Overhead	70%(L2+L3)	-	.87	51.1
12.	Pack & Ship	1000 MH	3.90	.07	3.9
13.	Waste Disposal	1000 MH	3.90	.07	3.9
14.	Other Indirect	-	-	-	-
15.	TOTAL INDIRECT	-	-	$ 1.33	$ 77.8
16.	Depreciation	10% FC	-	16.61	980.0
17.	Taxes	2% FC	-	3.32	196.0
18.	Insurance	1% FC	-	1.66	98.0
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$ 21.59	$ 1,274.0
21.	TOTAL OPERATING	-	-	$ 58.74	$ 3,463.3
22.	Credits HP Steam	18.3 M Lb.	.50	9.15	539.9
23.	NET OPERATING	-	-	$ 49.59	$ 2,923.4

EXHIBIT 10-26

METHANE REDUCTION OF SO_2
CASE 4b
WEST PROCESS
GAS 2B
SUMMATION OF MAJOR EQUIPMENT

1.	Cyclone collector system - multiple steel cyclone collectors with ash gates.	$ 40,000
2.	Precipitator - rod curtain design electrostatic precipitator, carbon steel construction for 70,000 SCFM 440°F gas.	$150,000
3.	Primary Blower - centrifugal gas blower to handle 70,000 SCFM gas at 440°F with a ΔP of 5 psi complete with 5,000 HP condensing steam turbine driver.	$300,000
4.	Waste Heat Boiler #1 - forced circulation water tube boiler with superheater with combustion chamber for methane-smelter gas reaction generating 137,000 lbs./hr. high pressure steam.	$300,000
5.	Reactor - Brick lined carbon steel reaction vessel 34 ft. diameter, 30 ft. high with 10 ft. alumina catalyst section supported on refractory grillwork.	$200,000
6.	Waste Heat Boiler #2 - water tube boiler with super heater generating 130,000 lbs./hr. high pressure superheated steam.	$330,000
7.	Intermediate Reactor - stainless steel conversion chamber 34 ft. diameter by 25 ft. high complete with 9,500 cu. ft. alumina catalyst supported on a cast iron grillwork, externally insulated.	$260,000
8.	Primary Condenser - low pressure steam generator of carbon steel construction designed to condense sulfur from 78,500 SCFM gas at 894°F at a rate of 22 short tons per day. Low pressure steam generation 55,000 lbs./hr.	$190,000
9.	Gas Preheater #1 - Shell and tube steam heater for 78,000 SCFM raising gas to 450°F for Claus #1 inlet. Heat duty 20 MM BTU/hr. using high pressure steam.	$ 90,000
10.	Claus Unit - Two stage Claus converter consisting of 45 ft. diameter carbon steel vertical vessel with two 7 ft. deep catalyst beds. Overall height 30 ft. Complete with 22,000 cubic feet alumina type catalyst and supports.	$370,000
11.	Process blower - Two stage centrifugal pressure blower for 4 PSIG P 78,000 SCFM at 260°F, complete with 4000 HP turbine driver.	$280,000

12.	Steam condensers - condensers for steam from turbine drivers for items 3 and 11 above, complete with auxiliaries.	$ 50,000
13.	Gas Preheater #2 - Shell and tube steam heater for 78,000 SCFM raising gas to 400°F for Claus #2 inlet. Heat duty 14 MM BTU/hr. using high pressure steam.	$ 70,000
14.	Claus condensers (2) - Shell and tube exchangers condensing sulfur from outlets of two stages of Claus converters. Boiler feedwater is preheated in the exchanger. Cost for two units.	$220,000
15.	Coalescer - carbon steel circulating liquid sulfur type coalescer of standard design, complete with auxiliaries including pump.	$ 40,000
16.	Tail Gas Heater - Shell and tube steam heater for tail gas prior to incineration, raising temperature to 700°F by exchange with superheated steam.	$100,000
17.	Incinerator and stack - carbon steel refractory-lined stack with lower section fitted as auxiliary heater to raise temperature of tail gas to 1200°F and discharge, complete with auxiliaries.	$200,000
18.	Sulfur Storage Tank - vertical 5,000 ton capacity with steam coils and external insulation.	$ 70,000
Total Purchased Equipment		$3,260,000
Total Installed Fixed Capital		$9,800,000

EXHIBIT 10-27
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: CH_4 CASE NO. 4c
PROCESS: Kulcsar GAS NO. 2B

Fixed Capital, Battery Limits Basis $9,400,000
Sulfur Production 180 NTD x 330 Days = 59,000 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	62.2 MCF	$.30 Avg.	$ 18.66	$ 1,100.0
1.	Total Raw Materials	-	-	$ 18.66	$ 1,100.0
2.	Direct Labor	18,000 MH	$3.50	1.07	63.0
3.	Supervision	1.	$10,000	.17	10.0
4.	Maintenance	5% FC	-	7.98	470.0
5.	Plant Supplies	10%(L2+L3)	-	.12	7.3
	Utilities:				
6. a)	Electricity	KWH	$.01	.70	41.3
b)	Cooling Water	MG	.05	2.33	137.5
c)	Boiler Feed Water	MG	.25	.28	16.5
d)	Fuel Gas	MCF	.30	2.13	125.7
7.	Total Utilities	-	-	5.44	321.0
8.	Other Direct (Catalyst)	9.5 Lbs.	.15	1.43	83.2
9.	TOTAL DIRECT	-	-	$ 34.87	$ 2,054.5
10.	Payroll Burden	30% L2	-	.32	18.9
11.	Plant Overhead	70%(L2+L3)	-	.87	51.1
12.	Pack & Ship	1000 MH	3.90	.07	3.9
13.	Waste Disposal	1000 MH	3.90	.07	3.9
14.	Other Indirect	-	-	-	-
15.	TOTAL INDIRECT	-	-	$ 1.33	$ 77.8
16.	Depreciation	10% FC	-	15.90	940.0
17.	Taxes	2% FC	-	3.18	188.0
18.	Insurance	1% FC	-	1.59	94.0
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$ 20.67	$ 1,222.0
21.	TOTAL OPERATING	-	-	$ 56.87	$ 3,354.3
22.	Credits HP Steam	14.5 M Lb.	.50	7.25	427.3
23.	NET OPERATING	-	-	$ 49.62	$ 2,927.0

EXHIBIT 10-28

METHANE REDUCTION OF SO$_2$
CASE 4c

KULCSAR PROCESS

GAS 2B

SUMMATION OF MAJOR EQUIPMENT

1.	Cyclone collector system - multiple steel cyclone collectors with ash gates.--	$ 40,000
2.	Precipitator - rod curtain design electrostatic precipitator, carbon steel construction for 70,000 SCFM 440°F gas.------	$150,000
3.	Primary Blower - centrifugal gas blower to handle 70,000 SCFM gas at 440°F with a ΔP of 5 psi complete with 5000 HP condensing steam turbine driver.--------------------------	$300,000
4.	Waste Heat Boiler #1 - forced circulation water tube boiler with superheater with combustion chamber for methane-smelter gas reaction generating 109,000 lb./hr. high pressure steam.--	$270,000
5.	Reactor - Brick lined carbon steel reaction vessel 22 ft. diameter, 15 ft. high with 2 ft. alumina catalyst section supported on refractory grillwork.-----------------------	$ 60,000
6.	Waste Heat Boiler #2 - water tube boiler with super heater generating 150,000 lb./hr. high pressure superheated steam.--	$350,000
7.	Intermediate Reactor - stainless steel conversion chamber 34 ft. diameter by 25 ft. high complete with 9500 cu. ft. alumina catalyst supported on a cast iron grillwork, externally insulated.--------------------------------------	$260,000
8.	Primary Condenser - low pressure steam generator of carbon steel construction designed to condense sulfur from 78,500 SCFM gas at 894°F at a rate of 22 short tons per day. Low pressure steam generation 60,000 lb./hr.--------------	$200,000
9.	Gas Preheater #1 - Shell and tube steam heater for 78,000 SCFM raising gas to 450°F for Claus #1 inlet. Heat duty 20 MM BTU/hr. using high pressure steam.------------------	$ 90,000
10.	Claus Unit - Two-stage Claus converter consisting of 45 ft. diameter carbon steel vertical vessel with two 7 ft. deep catalyst beds. Overall height 30 ft. Complete with 22,000 cubic feet alumina type catalyst and supports.------------	$370,000

EXH. 10-28 (Cont.)

11.	Process blower - Two-stage centrifugal pressure blower for 4 PSIG ΔP 78,000 SCFM at 260°F, complete with 4000 HP turbine drive.--	$280,000
12.	Steam condensers - condensers for steam from turbine drivers for items 3 and 11 above, complete with auxiliaries.------	$ 50,000
13.	Gas Preheater #2 - Shell and tube steam heater for 78,000 SCFM raising gas to 400°F for Claus #2 inlet. Heat duty 14 MM BTU/hr. using high pressure steam.-----------------	$ 70,000
14.	Claus condensers (2) - Shell and tube exchangers condensing sulfur from outlets of two stages of Claus converters. Boiler feedwater is preheated in the exchanger. Cost for two units.---	$220,000
15.	Coalescer - carbon steel circulating liquid sulfur type coalescer of standard design, complete with auxiliaries including pump.--	$ 40,000
16.	Tail Gas Heater - shell and tube steam heater for tail gas prior to incineration, raising temperature to 700°F by exchange with superheated steam.------------------------	$100,000
17.	Incinerator and stack - carbon steel refractory lined stack with lower section fitted as auxiliary heater to raise temperature of tail gas to 1200°F and discharge, complete with auxiliaries.---------------------------------------	$200,000
18.	Sulfur Storage Tank - vertical 5000 ton capacity with steam coils and external insulation.---------------------------	$ 70,000

Total Purchased Equipment.-------------------------------------	$3,120,000
Total Installed Fixed Capital -------------------------------	$9,400,000

EXHIBIT 10-29a
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: H_2S CASE NO. 6
PROCESS: Low Temp. Claus GAS NO. 2C
Fixed Capital, Battery Limits Basis $7,700,000
Sulfur Production 385 NTD x 330 Days = 127,000 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	12.2 MCF	$.30 Avg.	$ 3.66	$ 465.0
1.	Total Raw Materials	-	-	$ 3.66	$ 465.0
2.	Direct Labor	18,000 MH	$3.50	.50	63.0
3.	Supervision	1.	$10,000	.08	10.0
4.	Maintenance	5% FC	-	3.03	385.0
5.	Plant Supplies	10%(L2+L3)	-	.06	7.3
	Utilities:				
6. a)	Electricity	281 KWH	$.01	2.81	357.0
b)	Cooling Water	4.86 MG	.05	.24	30.5
c)	Boiler Feed Water	1.27 MG	.25	.32	40.6
d)	Fuel Gas	25.4 MCF	.30	7.60	965.0
7.	Total Utilities	-	-	10.97	1,393.1
8.	Other Direct (Catalyst)	85 Lbs.	.13	11.00	1,400.0
9.	TOTAL DIRECT	-	-	$ 29.30	$ 3,723.3
10.	Payroll Burden	30% L2	-	.15	18.9
11.	Plant Overhead	70%(L2+L3)	-	.40	51.1
12.	Pack & Ship	1000 MH	3.90	.03	3.9
13.	Waste Disposal	1000 MH	3.90	.03	3.9
14.	Other Indirect	-	-	-	-
15.	TOTAL INDIRECT	-	-	$ .61	$ 77.8
16.	Depreciation	10% FC	-	6.05	770.0
17.	Taxes	2% FC	-	1.21	154.0
18.	Insurance	1% FC	-	.60	77.0
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$ 7.86	$ 1,001.0
21.	TOTAL OPERATING	-	-	$ 37.77	$ 4,802.1
22.	Credits HP & LP Steam	5.6/2.7 M Lb.	.50/.25	3.50	444.0
23.	NET OPERATING	-	-	$ 34.27	$ 4,358.1

EXHIBIT 10-29b
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT:	H_2S	CASE NO.	6 (Factored)
PROCESS:	Low Temp. Claus	GAS NO.	2C

Fixed Capital, Battery Limits Basis $5,100,000
Sulfur Production 192 NTD x 330 Days = 63,500 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	12.2 MCF	$.30 Avg.	$ 3.66	$ 232.5
1.	Total Raw Materials	-	-	$ 3.66	$ 232.5
2.	Direct Labor	18,000 MH	$3.50	1.00	63.0
3.	Supervision	1.	$10,000	.16	10.0
4.	Maintenance	5% FC	-	4.01	255.0
5.	Plant Supplies	10%(L2+L3)	-	.12	7.3
	Utilities:				
6. a)	Electricity	281 KWH	$.01	2.81	178.5
b)	Cooling Water	4.86 MG	.05	.24	15.2
c)	Boiler Feed Water	1.27 MG	.25	.32	20.3
d)	Fuel Gas	25.4 MCF	.30	7.60	482.5
7.	Total Utilities	-	-	10.97	696.5
8.	Other Direct (Catalyst)	85 Lbs.	.13	11.00	697.5
9.	TOTAL DIRECT	-	-	$ 30.92	$ 1,961.8
10.	Payroll Burden	30% L2	-	.30	18.9
11.	Plant Overhead	70%(L2+L3)	-	.80	51.1
12.	Pack & Ship	1000 MH	3.90	.06	3.9
13.	Waste Disposal	1000 MH	3.90	.06	3.9
14.	Other Indirect	-	-	-	-
15.	TOTAL INDIRECT	-	-	$ 1.22	$ 77.8
16.	Depreciation	10% FC	-	8.03	510.0
17.	Taxes	2% FC	-	1.61	102.0
18.	Insurance	1% FC	-	.80	51.0
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$ 10.44	$ 663.0
21.	TOTAL OPERATING	-	-	$ 42.58	$ 2,702.6
22.	Credits HP & LP Steam	5.6/2.7 M Lb.	.50/.25	3.50	222.0
23.	NET OPERATING	-	-	$ 39.08	$ 2,480.6

EXHIBIT 10-30

LOW TEMPERATURE CLAUS
H_2S REDUCTION OF SO_2
2C SMELTER GAS
CASE 6
SUMMATION OF MAJOR EQUIPMENT

1.	Methane Reformer - direct gas fired box type reforming furnace designed to reform 3100 SCFM methane at 1570°F, 150 PSIG outlet incorporating charge preheat coils and steam generation coils in convection section. Thermal efficiency 85% including economizer coils. Steam generation 36000 lb./hr. at 250 PSIG. Complete with twin carbon bed desulfurizers.	$400,000
2.	Hydrogen Sulfide Synthesis Converter - comprising a 7 ft. diameter x 12 ft. high packed section for gas cooling and reaction and a 15 ft. diameter by 10 ft. high reaction section having two catalyst sections 2 ft. deep. Complete with 16 tons of catalyst.	$ 55,000
3.	Hydrogen Sulfide Reactor Intercooler - 250 PSIG steam generator used for interstage cooling of reaction stream, carbon steel construction.	$ 25,000
4.	Gas Cooler - water cooled shell and tube carbon steel heat exchanger suitable for H_2S service, 30 MM BTU/hr.	$ 35,000
5.	Liquid Sulfur Cooler - 150 PSIG steam generator used for cooling circulating sulfur stream. Carbon steel construction. 16000 lb./hr. steam generation,	$ 16,000
6.	Sulfur Pump Tank - carbon steel sulfur vertical hold tank with internal steam coils, 100 tons capacity, insulated.	$ 8,000
7.	Sulfur Circulation Pump - 250 GPM steam jacketed pump complete with 30 HP motor and variable speed drive.	$ 3,000
8.	Low Temperature Claus Converter - moving catalyst bed design. Chamber is steel, 33 ft. diameter x 35 ft. high straight side having two catalyst zones with internal downcomers and gas distribution system. Approximately 90% of gas flow is assumed to bypass through interzone water injection chamber to control bed temperature rise. H_2S is added at interstage point to prevent excessive temperature rise in bed. Including precooler, intercooler, catalyst and auxiliaries.	$300,000

9. Catalyst Regenerator - indirect heated moving bed design comprising 30 ft. diameter x 60 ft. high carbon steel shell with 760 - 6" tubes 37' long through which catalyst passes countercurrent to an inert or reducing circulation gas stream. Catalyst is distributed above the upper tube sheet with a rake or similar mechanical distributor. Heat for sulfur stripping is provided by condensing sulfur vapor on external tube surfaces. Including catalyst and auxiliaries.------- $500,000

10. Catalyst Cooler - Carbon steel 21 ft. diameter by 20 ft. high shell with air distribution/catalyst support grates. Tempered air cools 100 NT per hour catalyst from 790°F. to 260°F. Complete with catalyst, catalyst handling equipment, air tempering coils, air blower with 200 HP motor and auxiliaries.-- $135,000

11. Regenerator Gas Exchanger - shell and tube carbon steel heat exchanger designed to reheat circulating stripping gas stream to 700°F. by exchange against regenerator overhead stream. $ 10,000

12. Sulfur Coolers and Condensers - 35 PSIG steam generators designed to condense and recover 1200 NT/day sulfur from regenerator gas stripping system. Carbon steel construction. --- $200,000

13. Regenerator Circulation Blower - centrifugal blower of forged steel casing construction suitable for service on H_2S or nitrogen with P of 8 PSI complete with 800 HP motor.--- $130,000

14. Sulfur Boiler - firetube sulfur boiler with economizer to vaporize sulfur for heat transfer medium to regenerator. Total heat release 150 MM BTU/hr. Economizer heat recovery 40 MM BTU/hr. Thermal efficiency including economizer 75%. Including auxiliaries.---------------------------------- $130,000

15. Incinerator and stack - 150 foot stack with gas fired incinerator designed to release purified smelter gas stream and sulfur boiler flue gas at a temperature of 800°F.----- $350,000

16. Smelter Gas Compressor - single stage centrifugal compressor to handle 210,000 SCFM at an inlet temperature of 450°F. with a ΔP of 2 PSI. Equipped with a 4000 HP motor.------- $200,000

17. Sulfur Storage Tank - vertical steel tank 5000 ton capacity with steam coils and external insulation.---------------- $ 70,000

Total Purchased Equipment.----------------------------------- $2,567,000

Fixed Capital Cost -------------------------------------- $7,700,000

EXHIBIT 10-31b
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: H_2S CASE NO. 2R (Factored)
PROCESS: Low Temp. Claus GAS NO. 3
Fixed Capital, Battery Limits Basis $ 11,500,000
Sulfur Production 192 NTD x 330 Days = 63,500 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	12.4 MCF	$.30 Avg.	$ 3.72	$ 236.0
1.	Total Raw Materials	-	-	$ 3.72	$ 236.0
2.	Direct Labor	27,000 MH	$3.50	1.49	94.5
3.	Supervision	1.	$10,000	.23	15.0
4.	Maintenance	5% FC	-	9.05	575.0
5.	Plant Supplies	10%(L2+L3)	-	.17	11.0
	Utilities:				
6. a)	Electricity	2,580 KWH	$.006	15.45	980.0
b)	Cooling Water	0.46 MG	.05	.02	1.3
c)	Boiler Feed Water	1.40 MG	.25	.35	22.2
d)	Fuel Gas	18.7 MCF	.30	5.62	357.0
7.	Total Utilities	-	-	21.44	1,360.5
8.	Other Direct (Catalyst)	77 lbs.	.13	10.00	635.0
9.	TOTAL DIRECT	-	-	$46.10	$2,927.0
10.	Payroll Burden	30% L2	-	.45	28.4
11.	Plant Overhead	70%(L2+L3)	-	1.21	76.7
12.	Pack & Ship	1000 MH	3.90	.06	3.9
13.	Waste Disposal	1000 MH	3.90	.06	3.9
14.	Other Indirect	-	-		
15.	TOTAL INDIRECT	-	-	$ 1.78	$ 112.9
16.	Depreciation	10% FC	-	18.12	1,150.0
17.	Taxes	2% FC	-	3.62	230.0
18.	Insurance	1% FC	-	1.81	115.0
19.	Other Fixed	-	-		
20.	TOTAL FIXED	-	-	$23.55	$1,495.0
21.	TOTAL OPERATING	-	-	$71.43	$4,534.9
22.	Credits L.P. Steam	8.6 M Lbs	.25	2.15	136.2
23.	NET OPERATING	-	-	$69.28	$4,398.7

EXHIBIT 11-1

COSTS OF GAS

(From Chemical Week Survey, 10/5/68)

	Statewide as Burned ¢/MM Btu	Wellhead Nat Gas Prices ¢/MCF	Indust. Nat Gas Prices ¢/MCF
California	30.8	31.0	-
Los Angeles	-	-	34.9
San Francisco	-	-	-
Louisiana	19.9	18.3	-
Baton Rouge	-	-	23.5
Michigan	49.8	25.2	-
Detroit	-	-	53.0
Missouri	22.9	-	-
St. Louis	-	-	30.3
New York	36.5	31.0	-
Buffalo	-	-	60.4
N.Y. City	-	-	
Ohio	33.8	23.7	
Cincinnati	-	-	53.1
Cleveland	-	-	
Pennsylvania	31.1	28.4	-
Pittsburgh	-	-	51.9
Philadelphia			
Texas	19.8	13.0	-
Houston	-	-	19.2
Illinois	51.9	11.9	
Chicago	-	-	40.8

Exhibit 11-1 (Cont'd)

	Statewide as Burned ¢/MM Btu	Wellhead Nat Gas Prices ¢/MCF	Indust. Nat Gas Prices ¢/MCF
Phoenix, Ariz.	28.8	13.8	33.7
Albuquerque, N.M.	22.3	12.5	19.7
Idaho Falls, Idaho	-	-	37.8
Cheyenne, Wyo.	-	14.5	18.2
Charleston, W. Va.	40.9	23.6	40.5
Butte, Montana	19.7	8.3	27.1
Pueblo, Colo.	21.6	13.0	23.6
Louisville, Ky.	21.5	23.7	37.7

Take for high - the avg. of gas in Industrial Centers to Industrial Users = 41¢

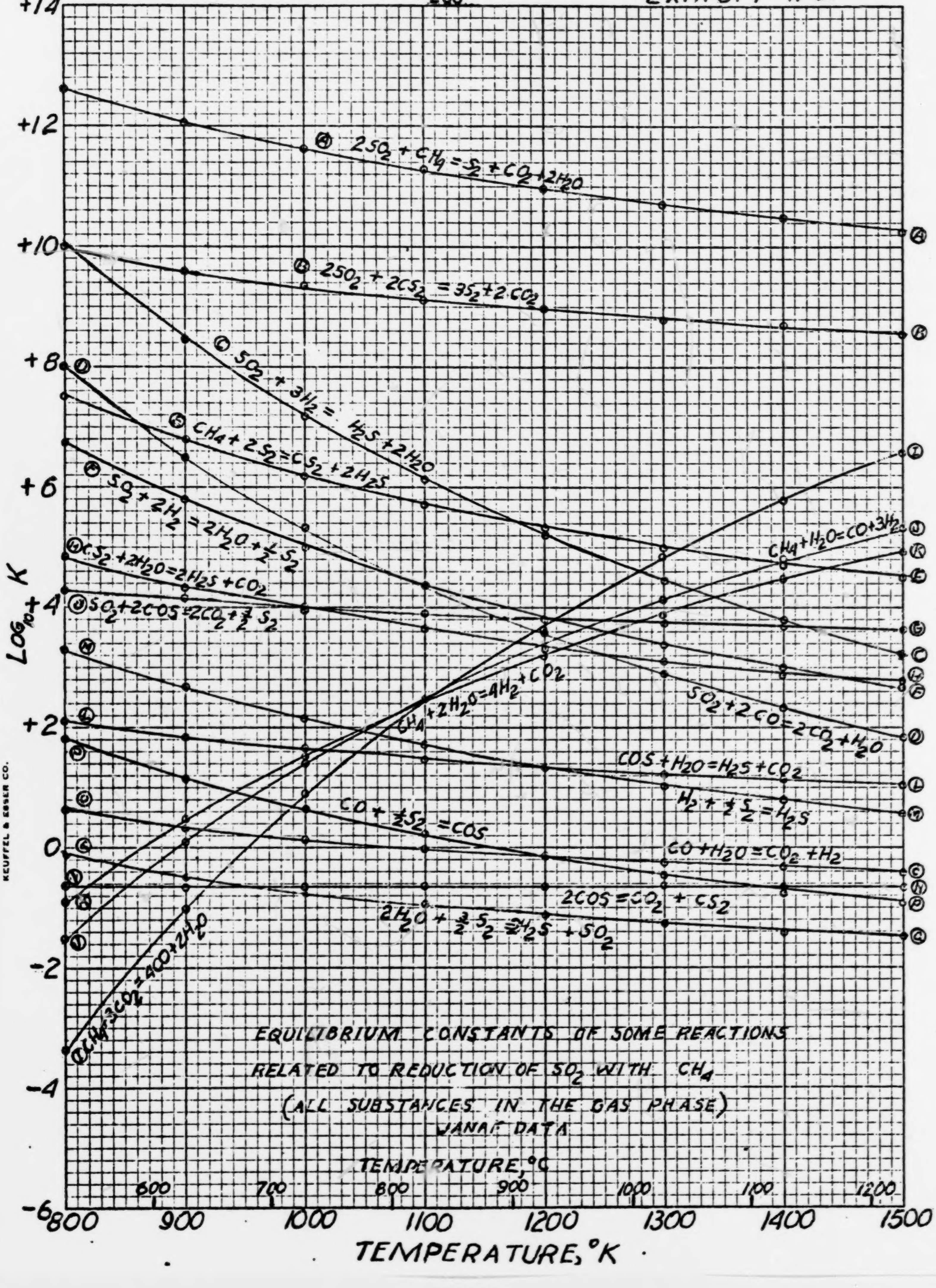
Take for low - overall avg. Wellhead gas price = 19¢

Take for avg. - Statewide avg. in Indust. States at 33¢
" " " Mining States at 25.8¢
This gives extra weight to mining areas 2 / 58.8 = 29.4¢
where metallurgical gases would be found

~~Reduces likelihood of dealing~~
~~power plant exits in Industrial Centers~~

Use 30¢

EXHIBIT 11-2



EQUILIBRIUM CONSTANTS OF SOME REACTIONS RELATED TO REDUCTION OF SO_2 WITH CH_4

(ALL SUBSTANCES IN THE GAS PHASE)

JANAF DATA

EXHIBIT 11-3

NOT REPRODUCIBLE

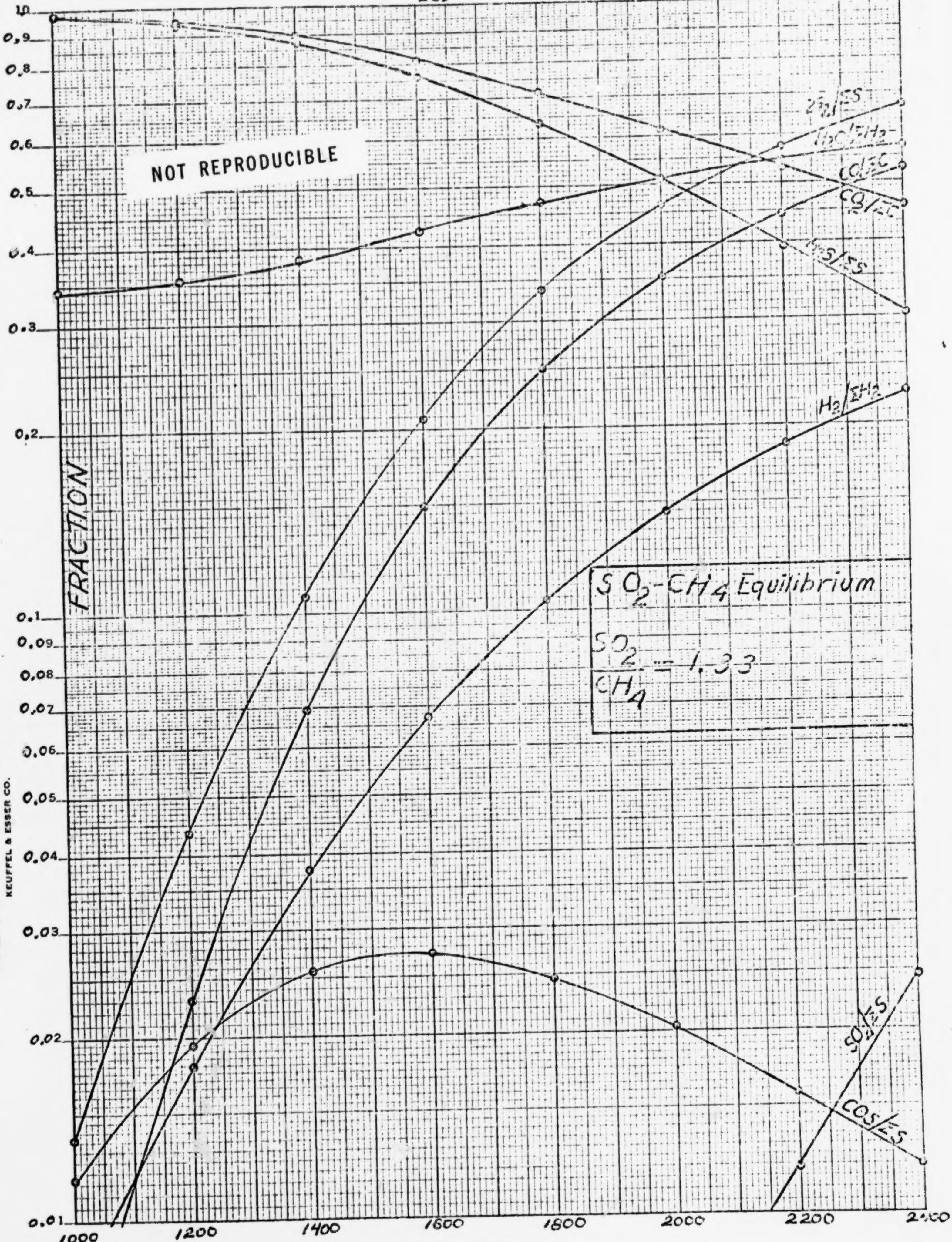


EXHIBIT 11-4

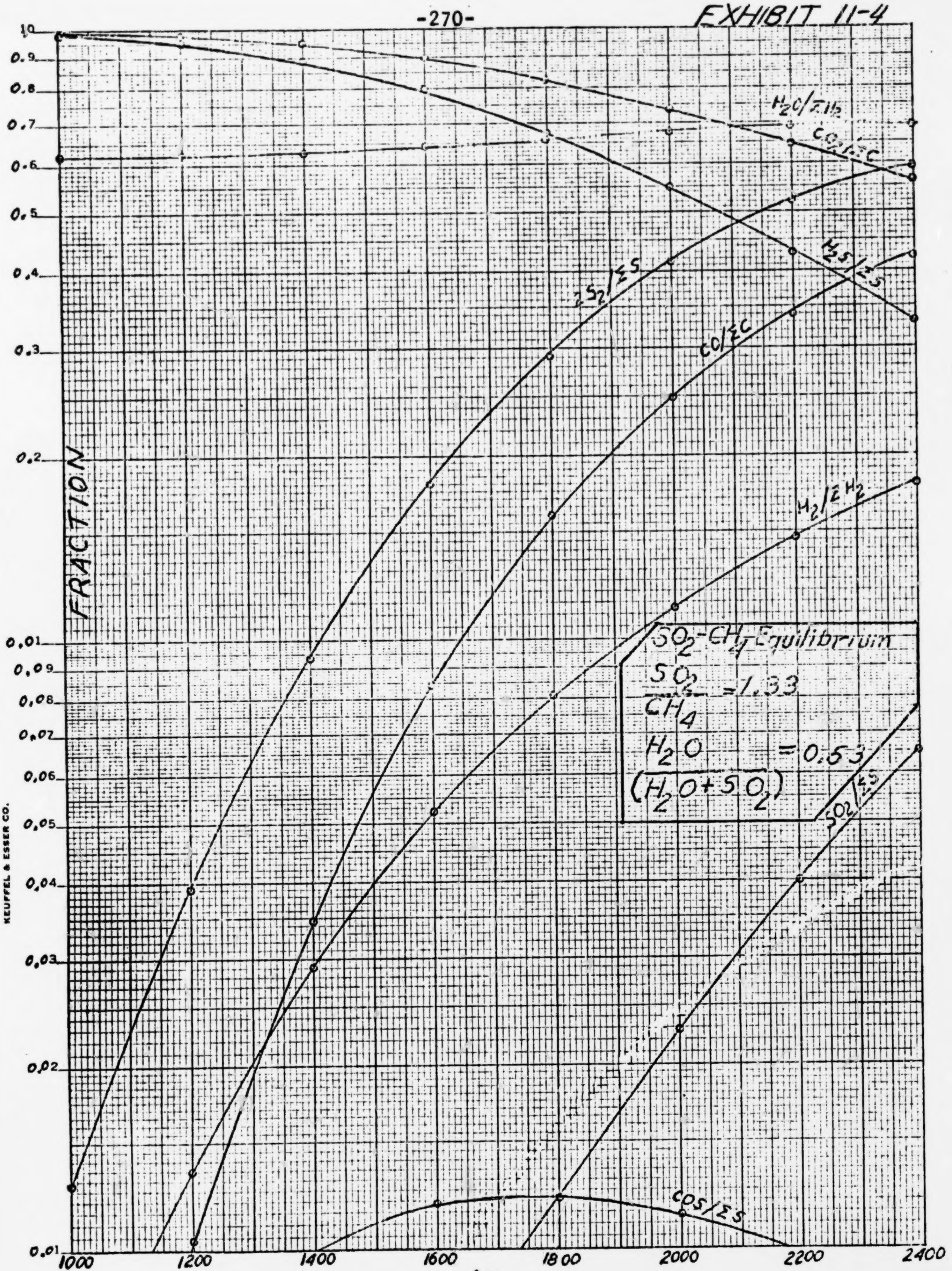

1.0
0.9
0.8
0.7
0.6
0.5
0.4
0.3
0.2
0.01
0.09
0.08
0.07
0.06
0.05
0.04
0.03
0.02
0.01
FRACTION
KEUFFEL & ESSER CO.
1000
1200
1400
1600
1800
2000
2200
2400
°F
$H_2O/\Sigma H_2$
$CO_2/\Sigma C$
$H_2S/\Sigma S$
$\frac{1}{2}S_2/\Sigma S$
$CO/\Sigma C$
$H_2/\Sigma H_2$
$SO_2/\Sigma S$
$COS/\Sigma S$
SO_2-CH_4 Equilibrium
$\frac{SO_2}{CH_4} = 1.33$
$\frac{H_2O}{(H_2O+SO_2)} = 0.53$

EXHIBIT 11-5a

COMPARISON OF KULCSAR'S EXPERIMENTAL COMPOSITION WITH EQUILIBRIUM COMPOSITIONS

REACTION TEMP	1100°C(2012°F)		1000°C (1832°F)					900°C(1652°F)				
	FEED RATIO 3.1		FEED RATIO 2.6			FEED RATIO 3.1		FEED RATIO 2.6			FEED RATIO 3.1	
	EQUILIB	SVH 6674	EQUILIB.	SVH 148	SVH 584	EQUILIB.	SVH 6674	EQUILIB.	SVH 148	SVH 58	EQUILIB.	SVH 6674
(Note 1) COMPONENT												
SO_2	30.74	27.39	21.60	21.86	25.48	30.42	37.35	21.57	20.82	34.97	30.41	42.99
CH_4	0	0.83	0	1.51	0.21	0	1.76	0	1.07	3.49	0	2.49
S_2	20.07	22.13	21.55	23.19	21.96	19.88	17.66	20.98	23.82	15.70	19.45	13.25
S_6	0	0	0	0	0	0	0	0	0	0	0	0
S_8	0	0	0	0	0	0	0	0	0	0	0	0
H_2S	4.54	1.34	7.27	3.96	1.86	5.26	2.03	8.47	3.65	2.69	6.15	3.03
CS_2	0	0	0	0	0	0	0.03	0	0	0.12	0	0.84
COS	0.16	0.57	0.22	0.11	0.92	0.15	0.81	0.20	0.09	2.89	0.14	1.40
CO	1.65	0.20	1.20	0.20	0.44	0.87		0.54	0.17	0.53	0.40	0.70
CO_2	22.59	22.85	26.38	25.93	26.35	23.37	21.84	27.05	27.90	20.87	23.86	19.02
H_2O	42.73	45.90	47.03	48.50	39.25	42.58	31.92	46.40	49.68	29.3[illegible]	42.11	25.06
H_2	1.53	0	1.30	0	14.13	0.96	11.43	0.73	0	16.79	0.54	15.85
TOTAL	124.01	121.21	126.55	125.26	130.6	123.49	124.83	125.94	127.20	127.37	123.06	124.63

REACTION TEMP	800°C (1472°F)				700°C (1292°F)				650°C (1202°F)		600°C (1112°F)	
	FEED RATIO 2.6		FEED RATIO 3.1		FEED RATIO 2.6		FEED RATIO 3.1		FEED RATIO 2.6		FEED RATIO 3.1	
COMPONENT (1)	EQUILIB.	SVH = 148	EQUILIB.	SVH= 6674	EQUILIB	SVH= 584	EQUILIB	SVH 6674	EQUILIB	SVH 584	EQUILIB	SVH= 6674
SO_2	21.92	25.66	30.70	53.21	22.63	43.15	31.29	70.38	23.11	53.50	32.15	73.51
CH_4	0	2.11	0	11.08	0	10.79	0	19.02	0	16.96	0	22.45
S_2	20.11	19.70	18.76	6.97	18.87	9.58	17.74	0	18.03	6.28	15.80	0
S_6	0	0	0	0	0	0	0	0	0.01	0	0.08	0
S_8	0	0	0	0	0	0	0	0	0.01	0	0.09	0
H_2S	9.87	1.08	7.24	4.58	11.68	3.76	8.71	0.13	12.77	3.24	10.55	0
CS_2	0	0	0	1.01	0	0.03	0	0	0	0	0.00	0
COS	0.17	0.76	0.12	1.85	0.14	6.06	0.10	5.09	0.13	2.93	0.08	2.02
CO	0.21	0.19	0.15	0.29	0.06	0.38	0.05	0.35	0.03	0.25	0.01	0
CO_2	27.41	22.65	24.12	10.24	27.59	10.64	24.25	0	27.64	7.79	24.31	0
H_2O	45.43	36.25	41.28	22.18	43.75	30.47	39.97	4.99	42.74	18.69	38.21	2.02
H_2	0.37	9.86	0.27	0	0.16	0	0.12	5.76	0.10	0	0.04	2.02
TOTAL	125.49	118.26	122.64	111.41	124.88	114.86	122.23	105.72	124.57	109.64	121.32	102.02

NOTES: (1) QUANTITIES IN MOLES ARE BASED ON 100 MOLES OF SO_2 AND REQUISITE MOLES OF CH_4, AS DETERMINED FROM FEED RATIO (MOLES OF SO_2/MOLE OF CH_4).

(2) ANALYTICAL DATA FROM KULCSAR (Oluj. TRANSYLVANIA UNIVERSITATE) STUDIA UNIVERSITATUM VICTOR BABES ET BOLYAI, SERIES I, VOL. 3, pp. 207-221 (1958) ELEMENTAL MATERIAL BALANCES SHOWED THE APPARENT PRESENCE OF FREE OXYGEN. THIS OXYGEN WAS CONVERTED BACK TO SO_2 BY THE REACTION:

$$S_2 + 2\,O_2 \longrightarrow 2\,SO_2$$

ICD-288 (8-67)

Averbukh's Data on Methane Reduction

Table 2

Results of reduction of 100% sulfur dioxide with methane

Space velocity, hr.$^{-1}$	Reaction time, sec.	Distribution of sulfur for SO_2/CH_4 = 1.9%: S_2	SO_2	H_2S	COS	CS_2	Conversion of SO_2, %	Distribution of sulfur for SO_2/CH_4 = 1.27%: S_2	SO_2	H_2S	COS	CS_2	Conversion of SO_2, %
Temperature: 900° C													
58	14.4	41.1	17.0	17.6	0	21.3	83.0	28.3	15.2	21.4	0	25.1	84.8
171	4.9	7.4	75.9	9.4	3.9	3.4	24.1	13.7	58.8	10.3	0.3	16.9	41.2
245	3.5	2.0	72.6	14.6	0	10.8	27.4	24.8	41.2	16.2	0.1	17.7	58.8
458	1.8	1.5	80.4	5.5	0	2.6	9.6	0.1	83.5	12.6	0	3.8	16.5
657	1.3	0.8	82.8	14.4	0	2.0	17.2	3.1	78.9	12.6	0	5.4	21.1
1870	0.44	0	85.2	14.4	0.1	0.3	14.8	0	83.0	16.6	0.1	0.3	17.0
Temperature: 1100° C													
58	12.3	79.5	6.2	13.5	0.4	0.4	93.8	31.6	5.2	60.7	0.9	1.6	94.8
73.7	9.7	71.0	0	24.2	0	4.8	100	13.0	0	70.9	0	16.1	100
171	4.2	72.6	0.9	24.4	0	2.1	99.1	14.2	1.7	79.5	0	4.6	98.3
245	3.0	81.4	6,4	9.0	0	3.2	93.6	12.6	1.0	80.6	1.4	4.4	99,0
458	1.6	39.3	15,6	14.7	0	30.4	84.4	40.4	2.3	30.8	0	26.5	97.7
900	0.76	70.6	2,2	15.6	0	11.6	97.8	34.0	1.2	45.8	0	19.0	98.8
1870	0.28	10.7	51.9	13.4	1.7	22.3	48.1	12.4	43.6	17,4	0	26.6	56.4
Temperature: 1250° C													
58	11.1	77.0	9,1	12.4	0	1.5	90.9	22.0	0	75.6	0	2.1	100
73.7	8.8	65.7	0,8	30.6	0	2.9	99.2	12.7	1.0	78.3	1.6	6.4	99.0
245	2.6	77.7	8.9	13.1	0	0.3	94.1	23.0	6.5	69.9	0	0.6	93.5
458	1.4	80,8	6.6	8.1	0	4.5	93.4	18.7	2.1	74.8	0	4.4	97.9
960	0.67	78.7	0.9	19.0	0	1.4	99.1	15.6	1.2	80.3	0	2.9	98.8
1870	0.34	71.0	1.8	18.2	0	9.0	98.2	23.8	0.4	50.7	0	25.1	99.6

Table 3

Results of reduction of 10-40% sulfur dioxide with methane at 1250° C

Space velocity, hr.$^{-1}$	Reaction time, sec.	Distribution of sulfur for 10% SO_2, %: S_2	SO_2	H_2S	COS	CS_2	Conversion of SO_2, %	Distribution of sulfur for 20% SO_2, %: S_2	SO_2	H_2S	COS	CS_2	Conversion of SO_2, %	Distribution of sulfur for 40% SO_2, %: S_2	SO_2	H_2S	COS	CS_2	Conversion of SO_2, %
SO_2/CH_4 molar ratio: 1.3																			
50	12.9	52,6	0	44.9	1.4	1.1	100	43.9	0	51.8	0	4.3	100	56.3	0	40.2	2,8	0.7	100
84	7,7	49.6	38.4	11.4	0	0.6	61,6	47.5	0	48,0	0	4,5	100	46.2	0	48.3	0	5.5	100
245	2.6	41.4	0	55.8	0	3.1	100	36.3	0	58.4	0	5.3	100	—	—	—	—	—	—
283	2.3	—	—	—	—	—	—	47.9	0	47,7	0	4.4	100	53.9	0	42.4	1,0	3.6	100
1870	0.34	35.5	0	56,3	0	8.2	100	41.4	0	47.6	0	11.0	100	48.8	0	38.7	2,0	10.5	100
SO_2/CH_4 molar ratio: 1.45																			
66	9.9	58.3	0	39.9	0	1.8	100	49.4	0	45.9	1,7	3.0	100	—	—	—	—	—	—
245	2.6	56.0	0	42,5	0,2	1.3	100	34.8	0	61.9	0	3.3	100	55.2	0	41.4	0	3.4	100
506	1.3	66,9	0,4	29.4	0	3.3	99.6	60.9	0	34.8	0	4.3	100*	59.9	0	34.2	0	5,9	100
1870	0.34	55.1	0	40.7	0	4.2	100	48.0	0	40,3	0	11.7	100	—	—	—	—	—	—
SO_2/CH_4 molar ratio: 2																			
50	12.9	70.7	11,6	16.4	0.3	0,8	88.2	74.8	5,6	17.8	0	1.8	94.4	76.1	10.0	12.7	0	1.2	90.0
84	7,7	65.5	15,3	17.5	0	1.7	84.7	72.7	7,6	17.8	0	1.9	92.4	78.8	8.4	11.3	0	1.5	91.6
245	2.6	65.8	11.4	21.2	0	1.6	88.8	76.9	4,0	17.5	0	1.6	96.0	75.4	12.1	11.2	0	1,3	87,9**
420	1.6	—	—	—	—	—	—	71.5	13,6	13,5	0	1,4	86.4	78.4	9.8	9.7	0	2.1	90,2***
513	1.3	60,6	10,9	17.2	0	1.3	82,8	—	—	—	—	—	—	—	—	—	—	—	—
1870	0.34	58,4	21.4	16.9	0,4	2.9	78.0	68.6	1.7	23.7	0	6.0	98,3	66.1	12.5	18.1	0	3.3	87,5
2830	0.23	—	—	—	—	—	—	46.9	29.8	10.4	0	12.9	70,2	—	—	—	—	—	--

---

* $SO_2 : CH_4$ = 1.3.
** $SO_2 : CH_4$ ~ 2.2.
* $SO_2 : CH_4$ ~ 2.1.

EXHIBIT 11-6
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: H_2S CASE NO. 2B
PROCESS: Low Temp Claus- Switch Reactors GAS NO. 3

Fixed Capital, Battery Limits Basis $ 19,000,000
Sulfur Production 385 NTD x 330 Days = 127,000 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	12.4 MCF	$.30 Avg.	$ 3.72	$ 472.0
1.	Total Raw Materials	-	-	$ 3.72	$ 472.0
2.	Direct Labor	27,000 MH	$3.50	.74	94.5
3.	Supervision	-	$10,000	.12	15.0
4.	Maintenance	5% FC	-	7.47	950.0
5.	Plant Supplies	10%(L2+L3)	-	.09	11.0
	Utilities:				
6. a)	Electricity (20,000 HP)	1040 KWH	$.006	6.23	791.0
b)	Cooling Water (3,000 gpm)	11.25 MG	.05	.56	71.1
c)	Boiler Feed Water (70 M lbs/hr)	0.525 MG	.25	.13	16.5
d)	Fuel Gas	23.1MCF	.30	6.93	880.0
7.	Total Utilities	-	-	13.85	1,758.6
8.	Other Direct (Catalyst) (life 1 yr)	63.0 Lbs	.125	7.87	1,000.0
9.	TOTAL DIRECT	-	-	$ 33.86	$4,301.1
10,	Payroll Burden	30% L2	-	.22	28.4
11,	Plant Overhead	70%(L2+L3)	-	.61	76.7
12.	Pack & Ship	1000 MH	3.90	.03	3.9
13.	Waste Disposal	1000 MH	3.90	.03	3.9
14.	Other Indirect	-	-		
15.	TOTAL INDIRECT	-	-	$ .89	$ 112.9
16.	Depreciation	10% FC	-	14.95	1,900.0
17.	Taxes	2% FC	-	3.00	380.0
18.	Insurance	1% FC	-	1.50	190.0
19.	Other Fixed	-	-		
20.	TOTAL FIXED	-	-	$ 19.45	$2,470.0
21.	TOTAL OPERATING	-	-	$ 54.20	$6,884.0
22.	Credits (LP Steam)			-	-
23.	NET OPERATING	-	-	$ 54.20	$6,884.0

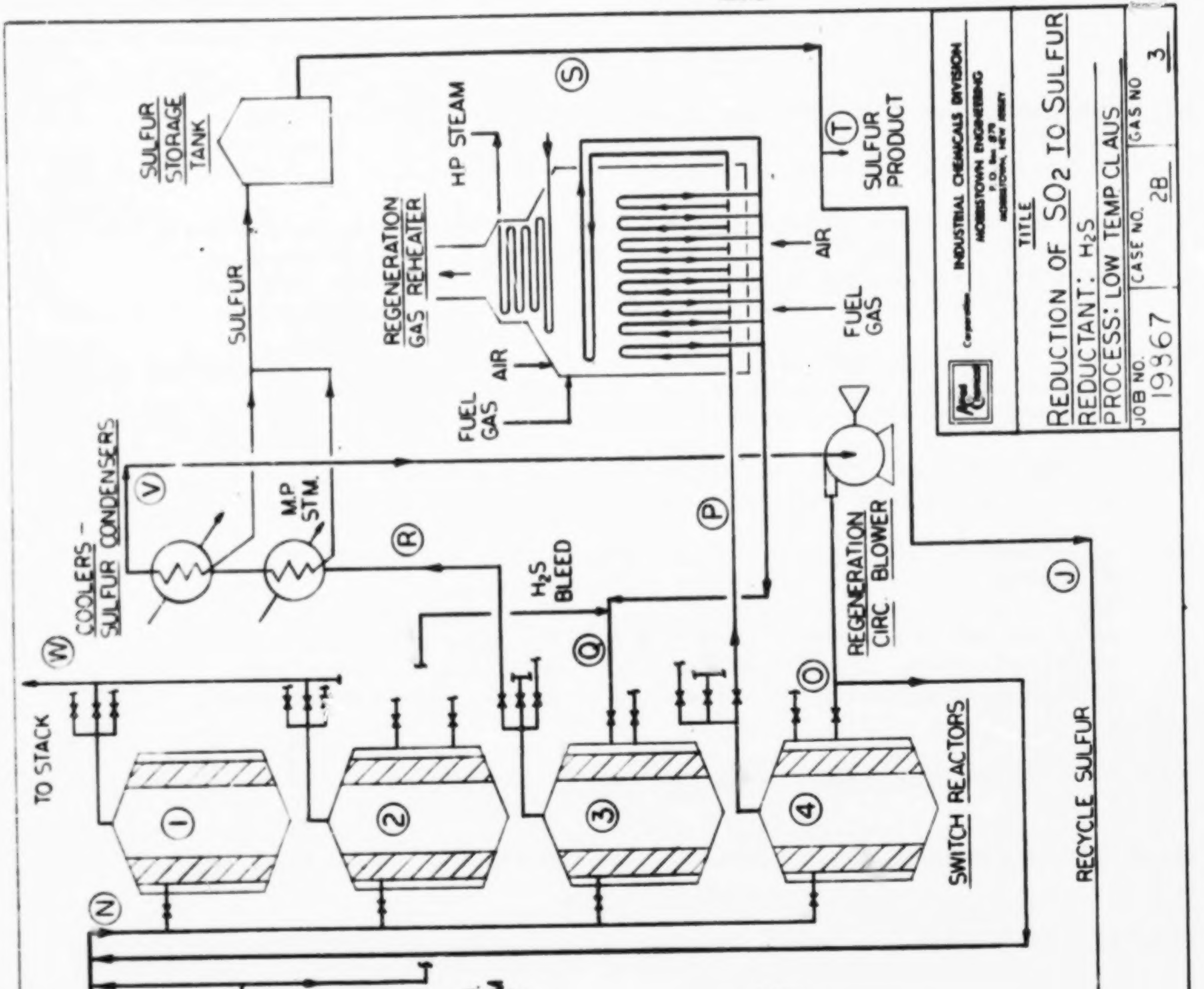
EXHIBIT 11-7

LOW TEMPERATURE CLAUS
H_2S REDUCTION OF SO_2
#3 STACK GAS
CASE 2B
SUMMATION OF MAJOR EQUIPMENT

1.	Methane Reformer - direct gas fired box type reforming furnace designed to reform 3600 SCFM methane at 1570°F, 150 PSIG outlet incorporating charge preheat coils and steam generation coils in convention section. Thermal efficiency 85% including economizer coils. Steam generation 40,000 lb./hr. at 250 PSIG. Complete with twin carbon bed desulfurizers.--------------	$425,000
2.	Hydrogen Sulfide Synthesis Converter - comprising a 7 ft. diameter x 12 ft. high packed section for gas cooling and reaction and a 14 ft. diameter by 3 ft. high reaction section having two catalyst beds 2 ft. deep. Complete with catalyst.--	$ 60,000
3.	Hydrogen Sulfide Reactor Intercooler - 150 PSIG steam generator used for interstage cooling of reaction stream, carbon steel construction, 22,000 lb./hr. steam generation.------------	$ 25,000
4.	Gas Cooler - water cooled shell and tube carbon steel heat exchanger suitable for H_2S service 34 MM BTU/hr.----------	$ 55,000
5.	Liquid Sulfur Cooler - 150 PSIG steam generator used for cooling circulating sulfur stream. Carbon steel construction. 18,000 lb./hr. steam generation.------------------------	$ 18,000
6.	Sulfur Pump Tank - carbon steel vertical sulfur hold tank with internal steam coils, 100 tons capacity, insulated.--	$ 8,000
7.	Sulfur Circulation Pump - 250 GPM steam jacketed pump complete with 30 HP motor and variable speed drive.----------------	$ 3,000
8.	Flue Gas Cooler - carbon steel 30 ft. diameter by 50 ft. high vertical duct section complete with spray system circulating pump, settler and auxiliaries.--------------------------	$ 76,000
9.	Flue Gas Valves - eight, 18 ft. diameter hydraulically operated goggle valves, enclosed design, twelve, 8 ft. diameter hot blast type gate valves, water cooled, hydraulically activated, complete with hydraulic system and controls, included as major equipment.---	$2,000,000
10.	Gas Booster Fan - single stage blowers for 2.4 million SCFM at 255°F with $P = 20$ inches W.C. carbon steel construction, complete with 18,000 H.P. motor.------------------------	$350,000

11.	Regeneration Circulation Blower - alloy steel hot gas blower for 212,000 SCFM at $P = 3$ psi, 300°F. Complete with 2000 HP turbine driver.	$400,000
12.	Regenerator Gas Reheater - double upfired box type furnace, comprising 65-12" O.D. radiant tubes and with economizer section generating high pressure steam. Maximum net heat duty 185 MM BTU/hr. radiant, 43 MM BTU/hr. convection.	$900,000
13.	Sulfur Coolers and Condensers - two stage sulfur condensing exchangers of carbon steel construction suitable for H_2S service.	$500,000
14.	Sulfur Storage Tank - vertical steel tank 2000 ton capacity with steam coils and external insulation.	$ 40,000
15.	Low Temperature Claus Switch Converters - four 50 ft. diameter by 50 ft. straight side converters per Exhibit 11- , aluminized carbon steel shell. Complete with perforated stainless steel partitions and with 1000 tons catalyst each. Includes catalyst handling systems and auxiliaries.	$1,800,000
	Total Purchased Equipment	$6,000,000
	Total Fixed Capital	$18,000,000
	Catalyst at $250/ton - 4000 Tons	$1,000,000
	Total Plant	$19,000,000

EXHIBIT II-8


SULFUR STORAGE TANK
SULFUR
HP STEAM
REGENERATION GAS REHEATER
AIR
FUEL GAS
AIR
FUEL GAS
Ⓢ
Ⓣ SULFUR PRODUCT
COOLERS - SULFUR CONDENSERS
Ⓥ
M.P. STM.
Ⓡ
H₂S BLEED
Ⓟ
Ⓠ
REGENERATION CIRC. BLOWER
Ⓞ
Ⓙ
Ⓦ
TO STACK
1
2
3
4
SWITCH REACTORS
RECYCLE SULFUR
Ⓝ
GAS BOOSTER BLOWER
REGEN. SYSTEM VENT
Ⓛ
GAS COOLER
LIQUID SULFUR COOLER
Ⓗ
Ⓖ
SULFUR CIRCULATION PUMP
Ⓤ
Ⓚ
Ⓓ
SULFUR PUMP TANK
FLUE GAS COOLER
Ⓕ
Ⓔ
H₂S SYNTHESIS CONVERTER
Ⓒ
REACTOR INTERCOOLER
METHANE REFORMER
WATER
FLUE GAS Ⓜ
Ⓐ CH₄
Ⓑ STEAM

INDUSTRIAL CHEMICALS DIVISION
MORRISTOWN ENGINEERING
P.O. Box 879
MORRISTOWN, NEW JERSEY

TITLE
REDUCTION OF SO_2 TO SULFUR
REDUCTANT: H_2S
PROCESS: LOW TEMP CLAUS

JOB NO.	CASE NO.	GAS NO
19367	2B	3

DRAWING NO. 37199

EXH 11-9

CASE 2.B

LOW TEMPERATURE CLAUS
2.35 MM SCFM #3 STACK GAS
SWITCH REACTOR SYSTEM

PROCESS POINTS	A	B	C	D	E	F	K	L	M	N	O	P	Q	R	V	W
TEMP. °C	25	205	898	300	625	300	540	125	149	107	135	273	500	260	135	120
SCFM	3560	10670	21200	23000	23970	23970	21580	21500	2.35MM	2.46MM	0.212MM	0.212MM	0.212MM	0.214MM	0.212MM	2.45MM
CFM @ T°C	3880	18700	91000	48290	78830	50310	64310	31380	3.63MM	3.42MM	0.317M	0.424M	0.914M	0.418M	0.317M	3.53MM
% CH_4	100.00		0.35	0.32	0.11	0.11										
% H_2O		100.00	29.03	26.78	19.03	19.03	16.75	16.80	7.25	10.61						11.18
% H_2S					35.69	35.69	64.98	65.18		0.570						0.0265
% CO_2			4.85	4.48	10.53	10.53	16.17	16.26	14.70	14.19						14.23
% COS					0.15	0.15	0.34	0.34								
% CO			11.57	10.68	3.96	3.96	0.11	0.11								
% S_2					11.48	11.48	0.34									
% H_2			54.20	50.00	19.05	19.05	1.31	1.31		0.01						0.01
% SO_2									0.30	0.287						0.0133
% O_2									2.80	2.68						2.68
% N_2									74.95	71.65	100.00	100.00	100.00	98.90	100.00	71.86
% S_6														0.33		
% S_8				7.74										0.77		

POINT	SULFUR DISTRIBUTION		
Ⓢ	FROM CATALYST	1293.9	NT/DAY
Ⓤ	FROM GAS COOLER	9.4	" "
	TOTAL S RECOVERY	1303.3	" "
Ⓙ	RECYCLE SULFUR	913.7	" "
Ⓣ	NET S RECOVERY	389.6	" "

SO_2 CONVERSION

EQUILIBRIUM 96.9 %

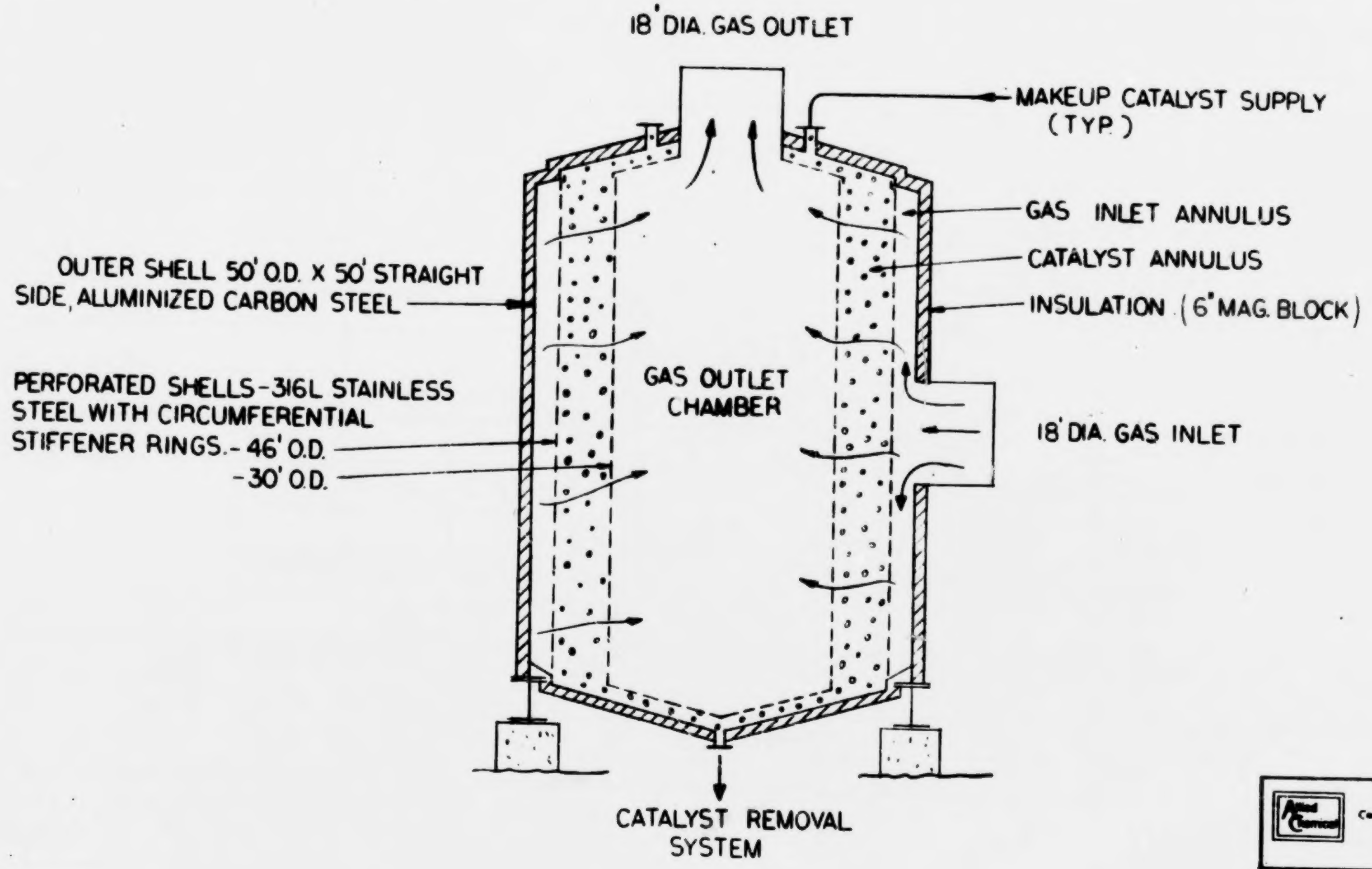
USED HERE 95.4 %

SULFUR YIELD 86.12 %

METHANE FACTOR 13160 SCF/TON S

EXHIBIT 11-9

EXHIBIT II-10


18' DIA. GAS OUTLET
MAKEUP CATALYST SUPPLY (TYP.)
GAS INLET ANNULUS
CATALYST ANNULUS
INSULATION (6" MAG. BLOCK)
18' DIA. GAS INLET
OUTER SHELL 50' O.D. X 50' STRAIGHT SIDE, ALUMINIZED CARBON STEEL
PERFORATED SHELLS-316L STAINLESS STEEL WITH CIRCUMFERENTIAL STIFFENER RINGS.- 46' O.D.
-30' O.D.
GAS OUTLET CHAMBER
CATALYST REMOVAL SYSTEM

Allied Chemical Corporation
INDUSTRIAL CHEMICALS DIVISION
MORRISTOWN ENGINEERING
P.O. Box 579
MORRISTOWN, NEW JERSEY

TITLE
REDUCTION OF SO_2 TO SULFUR
LOW TEMPERATURE CLAUS
SWITCH REACTOR

JOB NO.	CASE NO.	GAS NO.
19867	2B	3

DRAWING NO. 37201

EXHIBIT 11-11
PH 22-68-24

REDUCTION OF SO_2

REDUCTANT: H_2S - Low Temp. Claus CASE NO. 2A
PROCESS: Vert. Moving Bed Reactor GAS NO. 3

Fixed Capital, Battery Limits Basis $ 9,670,000
Sulfur Production 385 NTD x 330 Days = 127,000 NTY

Line No.	Cost Items	Used Per NT	Cost Per Unit	Per NT Sulfur	Per Yr. $M
	Methane	12.4 MCF	$.30 Avg.	$ 3.72	$ 472.0
1.	Total Raw Materials	-	-	$ 3.72	$ 472.0
2.	Direct Labor	27,000 MH	$3.50	.74	94.5
3.	Supervision	1.5	$10,000	.12	15.0
4.	Maintenance	5% FC	-	3.81	483.5
5.	Plant Supplies	10%(L2+L3)	-	.09	11.0
	Utilities:				
6. a)	Electricity	1,800 KWH	$.006	10.80	1,370.0
b)	Cooling Water	0.46 MG	.05	.02	2.5
c)	Boiler Feed Water	1.40 MG	.25	.35	44.5
d)	Fuel Gas	18.7 MCF	.30	5.62	715.0
7.	Total Utilities	-	-	16.79	2,132.0
8.	Other Direct (Catalyst)	77 lbs.	.13	10.00	1,270.0
9.	TOTAL DIRECT	-	-	$35.27	$4,478.0
10.	Payroll Burden	30% L2	-	.22	28.4
11.	Plant Overhead	70%(L2+L3)	-	.61	76.7
12.	Pack & Ship	1000 MH	3.90	.03	3.9
13.	Waste Disposal	1000 MH	3.90	.03	3.9
14.	Other Indirect	-	-	-	-
15.	TOTAL INDIRECT	-	-	$ .89	$ 112.9
16.	Depreciation	10% FC	-	7.62	967.0
17.	Taxes	2% FC	-	1.52	193.4
18.	Insurance	1% FC	-	.76	96.7
19.	Other Fixed	-	-	-	-
20.	TOTAL FIXED	-	-	$ 9.90	$ 1,257.1
21.	TOTAL OPERATING	-	-	$ 46.06	$ 5,848.0
22.	Credits L P Steam	8.6 M lb.	.25	2.15	273.0
23.	NET OPERATING	-	-	$ 43.91	$ 5,575.0

EXHIBIT 11-12

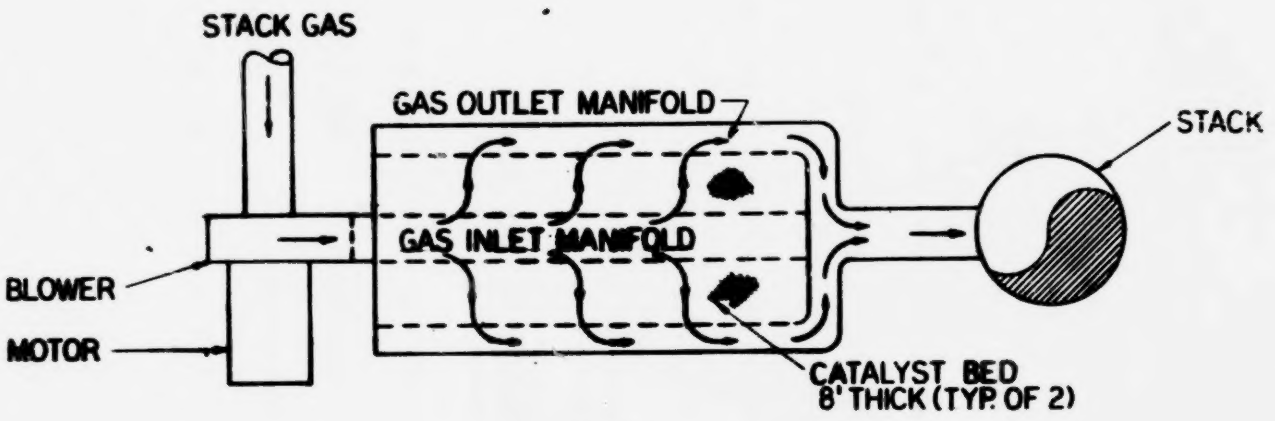
LOW TEMPERATURE CLAUS
H_2S REDUCTION OF SO_2
#3 STACK GAS
CASE 2A
SUMMATION OF MAJOR EQUIPMENT

1.	Methane Reformer - direct gas fired box type reforming furnace designed to reform 3600 SCFM methane at 1570°F., 150 PSIG outlet incorporating charge preheat coils and steam generation coils in convection section. Thermal efficiency 85% including economizer coils. Steam generation 40,000 lb./hr. at 250 PSIG. Complete with twin carbon bed desulfurizers.--------------	$425,000
2.	Hydrogen Sulfide Synthesis Converter - comprising a 7 ft. diameter x 12 ft. high packed section for gas cooling and reaction and a 14 ft. diameter by 8 ft. high reaction section having two catalyst beds 2 ft. deep. Complete with catalyst.--------	$ 60,000
3.	Hydrogen Sulfide Reactor Intercooler - 150 PSIG steam generator used for interstage cooling of reaction stream, carbon steel construction, 22,000 lb./hr. steam generation.------------	$ 25,000
4.	Gas Cooler - water cooled shell and tube carbon steel heat exchanger suitable for H_2S service 34 MM BTU/hr.------------	$ 55,000
5.	Liquid Sulfur Cooler - 150 PSIG steam generator used for cooling circulating sulfur stream. Carbon steel construction 18,000 lb./hr. steam generation.---------------------------------	$ 18,000
6.	Sulfur Pump Tank - carbon steel vertical sulfur hold tank with internal steam coils, 100 tons capacity, insulated.-------	$ 8,000
7.	Sulfur Circulation Pump - 250 GPM steam jacketed pump complete with 30 HP motor and variable speed drive.---------------	$ 3,000
8.	Flus Gas Stack - assumed existing.-----------------------	nil
9.	Low Temperature Claus Converter - Carbon Steel Vessel 35'x60'x100' comprising two catalyst chambers, inlet gas manifold, outlet gas manifold, catalyst hoppers and conveyors, catalyst transfer systems, associated ductwork and auxiliaries.--	$900,000
10.	Gas Booster Fan - single stage blower for 2.4 million SCFM at 255°F. with P = 20 inches W.C. carbon steel construction complete with 18,000 HP motor.----------------------------	$350,000

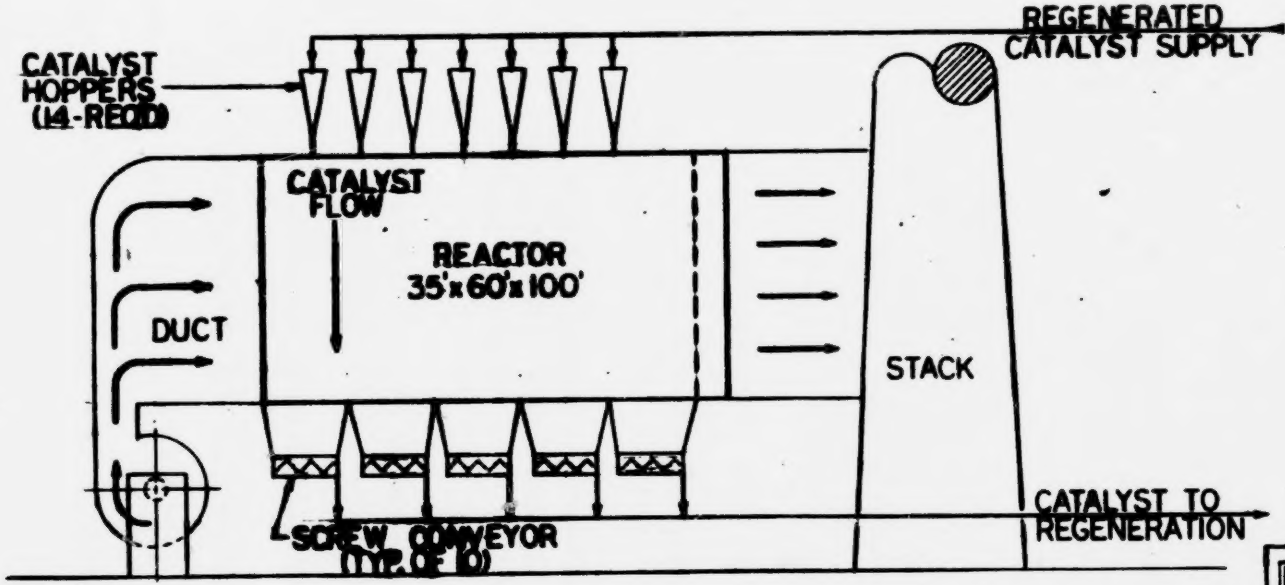
11.	Catalyst Regenerator - indirect heated moving bed design comprising 33 ft. diameter x 50 ft. high carbon steel shell with 820 - 6± tubes 37 ft. long through which catalyst passes countercurrent to an inert or reducing circulation gas stream. Catalyst is distributed above the upper tubesheet with a rake or similar mechanical distributor. Heat for sulfur stripping is provided by condensing sulfur vapor on external tube surfaces. Including auxiliaries.	$450,000
12.	Catalyst Cooler - Carbon steel 22 ft. diameter by 30 ft. high shell with air distribution/catalyst support gratings. Tempered air cools 220 NT per hour catalyst from 790°F. to 260°F. Complete with catalyst. Catalyst handling equipment, air tempering coils, air blower with 200 HP motor, and auxiliaries.	$130,000
13.	Regenerator Circulation Blower - centrifugal single stage blower of forged steel casing construction suitable for service on H_2S or nitrogen with P of 8 PSI complete with 800 HP motor.	$160,000
14.	Sulfur Boiler - firetube sulfur boiler with 150 PSIG steam economizer to vaporize sulfur for heat transfer medium to regenerator. Total heat release 172 MM BTU/hr. Economizer heat recovery 43 MM BTU/hr. steam generation 43,000 lb./hr. Thermal efficiency including economizer 75%. Including auxiliaries.	$120,000
15.	Regenerator Gas Exchanger - shell and tube carbon steel heat exchanger designed to reheat circulating stripping gas stream to 700°F. by exchange against regenerator overhead stream.	$ 10,000
16.	Sulfur Coolers and Condensers - 35 PSIG steam generators designed to condense and recover 1300 NT/day sulfur from regenerator gas stripping system. Carbon steel construction.	$170,000
17.	Sulfur Storage Tank - vertical steel tank 2,000 ton capacity with steam coils and external insulation.	$ 40,000

Total Purchased Equipment	$3,024,000
Total Fixed Capital, less catalyst	$9,075,000
Initial Catalyst Charge - 2275 tons	595,000
Total Fixed Capital	$9,670,000

EXHIBIT 11-13


STACK GAS
GAS OUTLET MANIFOLD
STACK
GAS INLET MANIFOLD
BLOWER
MOTOR
CATALYST BED
8' THICK (TYP. OF 2)

PLAN


REGENERATED
CATALYST SUPPLY
CATALYST
HOPPERS
(14-REQ'D)
CATALYST
FLOW
REACTOR
35'x60'x100'
DUCT
STACK
CATALYST TO
REGENERATION
SCREW CONVEYOR
(TYP. OF 10)

ELEVATION

Allied Chemical Corporation
INDUSTRIAL CHEMICALS DIVISION
MORRISTOWN ENGINEERING
P.O. Box 279
MORRISTOWN, NEW JERSEY

TITLE

REDUCTION OF SO_2 TO SULFUR
LOW TEMPERATURE CLAUS REACTOR
CONTINUOUS OPERATION

JOB NO.	CASE NO.	GAS NO.
19867	2A	3

DRAWING NO. 37200

16.0 BIBLIOGRAPHY

The entries are

I	Hydrogen Reduction
II	CO and COS Reduction
III	Mixtures of $CO + H_2$ and Other Reducing Gases
IV	Hot Carbon Reduction
V	Hydrocarbon Reduction
VI	H_2S - Normal Claus Reduction
VII	H_2S - Low Temperature Claus Reduction
VIII	Liquid Phase Reaction $H_2S + SO_2$
IX	Alkalized Alumina Reaction
X	Metallic Sulfides
XI	Miscellaneous Reduction Methods
XII	Review Articles on SO_2 Control
XIII	General References

In each classification, literature articles appear first, followed by a list of patents for that class. The entry numbers correspond with arbitrary numbers assigned during the complete literature search. Some items were discarded because they were of questionable relevance. The order of appearance is numerical for each class of article and each class of patent.

BIBLIOGRAPHY

I HYDROGEN REDUCTION:

52 Metallgesellschaft A.-G. Ger. 715,845, (Dec. 4, 1961).
SO_2 is passed through a bed of hot coal such that the maximum temperature is less than 800°C. The SO_2 is reduced almost completely. The exit gases are mixed with SO_2 to oxidize CO to CO_2 and COS to CO_2 and sulfur and passed over suitable catalysts at elevated temperatures.

60 Doumani, T. F. et al. Ind. Eng. Chem. 36, 329-32 (1944).
Waste gas mixed with hydrogen is passed thru a gel-type catalyst above 300°C and then over an activated alumina at 100° to 200°.

74 Deshmukh, G. S. and Lahiry, B. J. Indian Chem. Soc. 26, 257-61 (1949)
A mechanism is suggested for the interaction of sulfur dioxide and hydrogen under silent electric discharge: reaction products are S, H_2SO_3, H_2SO_4, SO_3, O_2, O_3, and H_2S.

172 Shakhtakhtinskii, G. B., and Guliev, A. I. Azerb. Khim. Zh. (1965) (2), 104-9.
Reduction with hydrogen was carried out in a quartz tube in a tubular oven at temperatures from 600-900° for a bauxite catalyst and 500-900° for an alunite catalyst, giving yields of 95-100% of sulfur.

202 Chedd, G. New Scientist 36 (#569) 281-3 (Nov. 1967).
Flue gases passed thru highly porous sodium aluminate react to form sodium sulfate, sodium sulfite, and Al oxide; then in a solids regenerator the mixture is reduced to H_2S.

342 Terres, Ernst, et al. Angew. Chem. 45, 164-71 (1932)
Reduction by CO or H_2 is practically quantitative up to 600°, decreasing with rising temperature, but amounting to 80% even at 800°.

351 Yushkevich, N. F., et al. J. Chem. Ind. (Moscow) 1933, No. 8, 50-8.
At 500-800° 40-60% yields of S result; the rest of the S goes into H_2S. The data of Terres et al. for equilibrium of the reaction are incorrect.

437 Berthelot, Marcellin P. E. Ann. Chim. Phys. (7) 14, 167-170 (1898)
If a mixture of hydrogen and SO_2 are ignited in a vessel open to the air, small quantities of H_2S and S are formed.

438 Berthelot, Marcellin P. E. Ann. Chim. Phys. (7) 14, 289-293 (1898)
At red heat SO_2 and hydrogen react to form S and H_2O; if the temperature is not too high, H_2S is formed.

470 Tomkinson, Margaret G. Compt. rend. 176, 35-6 (1923)
The action of Ni, NiS, CoS, and FeS as catalysts was reported.

558 Scharfenberg, Curt. PB Report 73552, Frames 5596-5607.
U.S. Dept. of Commerce, Office of Technical Services. Sept. 27, 1940. 12 pages.
Semi-plant scale experiments at 600-700° with a bauxite catalyst in a ceramic reactor gave 99.8% pure S.

PATENTS

62 Doumani, Thomas F. (to Union Oil Co. of Calif.) U.S. 2,361,825 (Oct. 31, 1944).
For reduction at 600-2400°F there is used a catalyst of Al_2O_3 on which a sulfide of Fe, Ni, or Co has been deposited.

100 Shiromoto, Yoshimitsu, and Hoshiba, Yukio (to Fuji Iron & Steel Co.) Japan. 6615 (Aug. 7, 1956).
Active H produced by thermal decomposition of hydrocarbon gases reduces SO_2 to H_2S.

231 Slaugh, Lynn H. (to Shell Oil Co.) U.S. 3,321,274 (May 23, 1967).
Reduction to H_2S took place at 100°C using a catalyst of Re_2S_7 in the presence of iso-PrOH.

304 Fischer, F. Brit. 282,634 (Dec. 24, 1926).
The catalyst is a precious metal; silverized asbestos may be used at 200-300°C.

329 Boswell, Maitland C. Can. 301,554 (July 1, 1930).
Reduction of 1 vol. SO_2 with 2 vol. H_2 is carried out at 600-800° over a V_2O_5 catalyst on asbestos, pumice or other inert material.

340 Boswell, Maitland C. Can 308,238 (Feb. 3, 1931).
A catalyst is prepared by treating an oxide of a metal of the Fe group with SO_2 and H_2; the catalyst is used for reduction of SO_2 at 300°.

346 Boswell, Maitland C. U. S. 1,880,741 (Oct. 4, 1932). An oxide of an Fe-group metal is treated with SO_2 and H_2 at a temperature sufficient to cause the oxide to take up S, a porous mass being formed; the material is used at 300° to effect reduction of SO_2 to S.

365 Bacon, Raymond F. Can. 332,225 (May 9, 1933). SO_2 and H_2 react at 250° in the presence of molten S and CuS to produce S.

371 Maier, Charles G. (one-fourth each to Reginald S. Dean and Henry A. Doerner) U. S. 1,941,702 (Jan. 2, 1934). A mixture of H (55% or more) with CO is used as the reducing agent at 700-900°.

400 Boswell, Maitland C. U. S. 2,026,819 (Jan. 7, 1936). Directions are given for preparing a catalyst from a hydrated sulfide of a metal of the Fe group.

493 Bacon, Raymond F. U. S. 1,840,076 (Jan. 5, 1932). SO_2 was subjected to the action of H in the presence of molten S.

II CO AND COS REDUCTION

1 Brownlie, David Steam Engr. 6, 408-10 (1937) The Boliden process, in which SO_2 is reduced with CO in the presence of a catalyst, is discussed as applied to flue gases.

2 Brownlie, David. Steam Engr. 6, 508-10 (1937) A review, including CO reduction.

3 Brownlie, David. Engr. of India 5, 70-3 (1937) Details are given for the I.C.I. and the Boliden processes.

9 Agren, Per. Svensk Papperstidn. 39, 470-82 (1936). In Sweden reduction is carried out at 400-600° using an Al-Fe oxides catalyst and a reducing gas (preparation described) consisting mostly of CO with small amounts of S compounds.

63 Visconti, Yvonne Stourze. Rev. quim. ind. (Rio de Janeiro) 14, No. 163, 22-6 (1945) Bauxite, impregnated with V salts, was used as catalyst in 197 tests carried out to obtain S from SO_2 and CO.

64 Visconti, Yvonne Stourdze. Anais assoc. quim. Brasil 4, 155-62 (1945).
See preceding reference.

105 Barea, Joaquin. Ion 18, 424-31 (1958)
Attempts to produce CS_2 from SO_2 with CO as reducing agent by using catalysts of Pt, Pd, or V_2O_4 impregnated in Kieselguhr or by using calcined bauxite gave negative results; large amounts of S were formed.

124 Krause, A. Bull, acad. polon sci., Ser. sci-Chim. 8, 439-40 (1960).
The reaction of COS with SO_2 using Fe_2O_3 as catalyst is studied.

173 Shakhtakhtinskii, G. B., and Guliev, A. I. Azerb. Khim. Zh. (1965) (6), 123-7
Reduction by CO at 500-900° was studied with an alunite catalyst at various flow rates and results were compared with those obtained with bauxite as catalyst.

221 Ryason, P. R., and Harkins, J. J. Air Poll. Control Assoc. 17 (12) 796-9 (Dec. 1967).
Simultaneous reduction of SO_2 and NO by catalyzed reaction with CO was studied; metals supported on alumina were the preferred catalysts.

234 Basmadzhieva, K., et al Gigiena i Sanit. 33, #4, 73-7 (1968).
Interaction of SO_2 and CO in the atmosphere is considered.

269 Lewis, G. N., and Lacey, W. N. J. Am. Chem. Soc. 37, 1976-83 (1915).
To calculate the free energy of formation of SO_2, the equilibrium between COS, CO and S was studied.

279 Ferguson, John B. Proc. Nat. Acad. Sci. 3,371-4 (1917)
The equilibrium between CO, CO_2, SO_2 and S was measured at 1000° and 1190° by a "semi-stream" method.

283 Ferguson, John B. J. Am. Chem. Soc. 40, 1626-44 (1918).
Details of the work reported in an earlier paper.

298 Yushevich, N. F., and Karzhavin, V. A. J. Chem. Ind. (Russia) 2, 474-8; 550-63 (1925).
The reduction by CO was studied at temperatures between 800° and 1180° with an FeS catalyst.

335 Yushkevich, N. F., and Karzhavin, V. A. J. Chem. Ind. (Moscow) 8, No. 1, 3-14; No. 2, 109-19 (1931).
The paper includes theoretical considerations, review of literature and descriptions of lab. and semi-plant experiments aimed at finding best conditions for production of S by reduction of SO_2. With CO as reductant the velocity of the reaction at temps. up to 600° is increased greatly by use of a catalyst consisting of 20-5% Fe_2O_3 and 80-75% Al_2O_3.

336 Yushkevich. N. F., and Karzhavin, V. A. J. Chem. Ind. (Moscow) 8, No. 11-12, 1053-8 (1931); No. 14, 1-11 91931)
Based on principles developed earlier, semi-factory-scale experiments are described.

428 Smith, Watson, and Hart, W. B. J. Soc. Chem. Ind. 5, 643-4 (1886).
A mixture of CO_2, CO, and SO_2 was passed through a red-hot glass tube containing pumice-stone fragments; S sublimed on the far end of the tube.

466 Stock, Alfred, and Seelig, Paul. Ber. 31, 681-94 (1919).
A study was made of the reactions that occur when COS is heated at different temperatures.

472 Stock, Alfred, et al. Ber. 57B, 719-35 (1924)
COS was heated at temperatures from 600° to 1100° and the decomposition products were analyzed for COS, CO, CO_2 and CS_2. The effects of Pt, glass, quartz, and charcoal as catalysts were determined.

528 Applebey, M. P. J. Soc. Chem. Ond. 56, 139-146T (1937).
The three stages of the Bolidens process are described: manufacture of reducing gas, reduction of SO_2, cooling of the gases and separation of the condensed S by electrostatic precipitation. Details of the Imperial Chemical Industries process are also described.

529 Avdeeva, A. V. J. Chem. Ind. (U.S.S.R.) 14, 1077-82 (1937)
Using a catalyst prepared from blast furnace slag containing about 55% Al_2O_3, the reaction of SO_2 with CS_2 and COS takes place satisfactorily, preferably at 400-425°.

579 Ishiwara, Torajiro, et al. Science Depts. Research Insts., Tohoku Univ., Ser. A, 3, 397-406, 407-12 (1951).

The effect of the following on the yields of S were determined: kind of catalyst, temperature, continuous use of catalyst, and relation between weight of catalyst and flow of gas.

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29 Bolidens Gruvaktiebolag (P.H.W. Agren, inventor). Swed. 91,742 (March 18, 1938).
An enriched gas, obtained with the aid of an adsorbent (active C or silical gel), is reduced with C or coke-oven gas; several modifications of the process are described.

30 Lindblad, Axel R. Ger. 660,641 (May 31, 1938).
SO_2 is reduced with generator gas over bauxite at 500°-1050° in the presence of vapors of alkali metal or compound thereof.

54 Lindblad, Axel R., and Herneryd, Fritz O. (to Bolidens Gruvaktiebolag) Ger. 721,053 (April 23, 1942).
S is produced from H_2S and SO_2 at 120-300° with a catalyst of cement or gypsum to which are added metallic sulfides.

240 Societe anon. Metallurgique, Procedes de Laval. Ger. 196,604 (Sept. 4, 1904).
Tangential introduction of CO into SO_2 promotes good mixing thereby aiding the reduction.

270 Hall, W. A. U. S. 1.134,846 (April 6, 1915)
Steam and a reducing gas flame is introduced into the upper part of the smelting chamber; the free S formed is kept out of contact with air until it cools sufficiently so that it will not be oxidized.

290 Shiomi, T. U. S. 1,359,114 (Nov. 16, 1920).
A mixture of SO_2, steam and CO is passed into a chamber filled with CaS, or an alk. metal sulfate, heated to a temperature of about 350-800°.

319 Benner, Raymond C., and Thompson, Alfred P. (to General Chemical Co.) U. S. 1,710,141 (April 23, 1929).
Gases containing COS together with an oxidizing gas such as O or SO_2 are brought into contact with activated bauxite.

391 Lindblad, Axel R. Brit. 426,850 (April 10, 1935).
Masses of aluminate or "Smelt" cement are used as the catalyst. Efficiency is increased by the addition of Fe or Mn compounds.

404 Lindblad, Axel R. U. S. 2,050,708 (Aug. 11, 1936).
The catalyst consists of cast cement relatively high in aluminiferous content and amounting to at least 25% of the total and relatively low in silica content.

488 Levy, S. I., and Leaver, C. British Patent 358,851 (Nov.5, 1931).
Sulfur was obtained by reaction of COS with SO_2 at 800°-1000° in contact with a refractory material impregnated with an iron compound.

516 Lindblad, Axel Rudolf. Norwegian 54,911 (Feb. 25, 1935).
As catalyst is used an aluminum cement with at least 25% Al_2O_3 alone or in mixture with Fe, Mn or compounds thereof.

553 Bolidens Gruvaktiebolag and Imperial Chemical Industries Ltd. Fr. 847,326 (Oct. 6, 1939).
Gases containing H_2S, COS and CS_2 are mixed with SO_2; the mixture is passed over a catalyst (Na_2O, Na_2S, Na_2CO_3) at a temperature over 350°.

640 Chevron Research Co. Fr. 1,523,685 (May 3, 1968).
SO_2 and oxides of nitrogen are removed from chimney gases by contacting the gases at at least 400°C with a Cu, Ag, Ni, Mo, Pd, or Co on alumina for at least 0.1 seconds in the presence of CO in at least 0.75 of stoichiometric.

III MIXTURES CO + H_2 AND OTHER REDUCING GASES:

89 Egorof, N. N. Desolfoazione dei gas di cokeria, del gas naturale e di altri gas combustibili. Milan, U. Hoepli, (1954). 256 pp.
Translation of a Russian book.

91 King, J. G. Coke and Gas 16, 437-48 (1954).
A review of the inaugural conference of the Institution of Fuel; deals with removal and recovery of S from fuels.

* 194 Shaktakhtinskii, G. B., and Guliev, A. I. Azerb. Neft. Khoz. 45 (3) 39-41 (1966).
Lab investigation at 500-900° with bauxite and alunite as catalysts.

* 531 Lepsoe, R. Ind. Eng. Chem. 30 pp 92-100(1937).
Equilib gas compositions on reduction of SO_2 by CS_2, H_2S, C, CO, H_2 and CH_4 are computed from thermodynamic data.

342 Terres, Ernst et al. Angew. Chem. 45, 164-71 (1932). Equilibrium reaction which take part in gas generators and S-purification processes are studied. Reduction by CO or H_2 is practically quantitative up to 600°.

539 Howat, D. D. Mines & Quarry Eng. 3, 91-8 (1938). Step 1: A portion of SO_2 is passed over coke in a specially designed producer, giving CO, H_2, small amounts of other gases, and sulfur vapor. Step 2: The latter mixture is used to reduce the rest of the SO_2 with the aid of a catalyst.

543 Palen, A. G. Paul Eng. Mining J. 139, 54-5 (Sept. 1938). Developments since 1932 at the Ronnskar smelter, which uses the Boliden process, are briefly outlined.

PATENTS

28 Metallgesellschaft A.-G. Fr. 825,289 (Feb. 28, 1938). Reduction is carried out by passage thru a bed of solid fuel in a furnace functioning as a generator, the temp. of the gas leaving the fuel being below 800°. Some SO_2 is added to the escaping gas to convert CO to CO_2 and to form S; the gaseous mixture is subjected to a catalytic treatment.

243 Smith, Paul S (to E. I. Du Pont de Nemours Powder Co.) U. S. 878,569 (Feb. 11, 1908). SO_2 is reacted with water gas.

247 Smith, Paul S. (to E. I. Du Pont de Nemours Powder Co.) U. S. 912,743 and 912,744 (Feb. 16, 1909). Water gas is mixed with SO_2 and the mixed gases are heated by producer gas.

348 Bacon, Raymond F. Brit. 378,668 (Aug. 18, 1932). Hot roaster gases are treated with water, producer or natural gas in a previously heated chamber; apparatus is described.

355 Bacon, Raymond F., and Bencowitz, Isaac (to Raymond F. Bacon) U. S. 1.917,234 (July 11, 1933). SO_2 from pyrite is reacted with natural gas or water gas; apparatus and process steps are described.

356 Bacon, Raymond F., and Bencowitz, Isaac (to Raymond F. Bacon) U. S. 1.917,685 (July 11, 1933).

Gases are introduced into a chamber at about 1000° and treated with a carbonaceous reducing gas; heat exchange procedures are described.

382 Metallgesellschaft A.-G. (Conway Freiherr von Girsewald and Hans Weidmann, investors) Ger. 594,464 (Mar. 17, 1934). A solid, liquid or gaseous reducing agent is reacted with SO_2 containing sufficient oxygen to maintain the reaction temperature by combustion of part of the reducing agent.

388 Lindblad, Axel R. Brit. 426,165 (Mar. 28, 1935) The reduction is effected in a chamber so constructed that the SO_2 is introduced in countercurrent to and is preheated by the reaction gases.

389 Lindblad, Axel R. Brit. 426,455 (Apr. 3, 1935) Reduction is effected in a chamber maintained at 500-1050°; the chamber is filled with contact material coated with Fe, Fe compound, or an alkali compound.

392 Lindblad, Axel R. Brit 427,999 (May 3, 1935) Gases evolved from smelting pyrite are brought in contact with a gaseous, liquid or powdered solid reducing agent; the resulting gases are passed thru a chamber filled with brick that has been coated with an alkali metal compound mixed or not with an Fe compound.

410 Lindblad, A. R. Swed. 85,183 (Jan. 9, 1936). A reaction chamber is so designed that the entering SO_2 is heated by the heat of reaction and moves in opposite direction to the reducing gas.

446 Smith, Paul S. (to E.I. duPont de Nemours Powder Co.) U.S. 945,111 (Jan. 4, 1910). Details are given for a system of generators and reports for the reduction of SO_2 by water gas.

447 Smith, Paul S. (to E.I. duPont de Nemours Powder Co.) U.S. 945,112 (Jan. 4, 1910). An apparatus is described for reduction with water gas in which system producer gas by its combustion supplies the heat required for the reaction.

* 592 Van Helden, H. J. A. (to Shell Internationale Research Maatschappy N. V.) Brit 1,116,129 (June 6, 1968). A process to reduce SO_2 to H_2S in which SO_2 and a reducing gas mixture contacts a catalyst of vanadium oxide on a carrier material.

469 Chemische Fabrik Thenania A.-G. and B. C. Stuer. Ger. 305,621 (Nov. 20, 1922)
Hydrated oxides of iron are used as catalysts; carriers containing water of hydration may be present.

471 Rhenania Verein Chemischer Fabriken A.-G. and B. G. Stuer Ger. 372,715 (March 31, 1923).
Metal sulfides are used as catalysts.

501 Bacon, Raymond F., and Hotchkiss, Henry T., Jr. (to Raymond F. Bacon) U. S. 1.917,687 (July 11, 1933).
In smelting sulfide ores, a reducing agent such as water gas or natural gas is introduced into the furnace charge above the combustion zone at a point where reaction with SO_2 can take place.

521 Lindblad, Azel R. U. S. 2,054,941 (Sept. 22, 1936).
Gases from the smelting of pyrite are mixed with producer gas and the mixture is passed into a chamber filled with hot contact material coated with an alkali metal compound.

535 Beckman, John W. U. S. 2,090,828 (Aug. 24, 1937).
Reduction is effect by CO, H, or CH_4 in contact with a porous lime at about 400°.

563 De Jahn, Fredrik W. Fr. 854,116 (April 5, 1940)
SO_2 is reduced to S by means of producer gas in presence of a catalyst consisting of a mixture of a metal of Group VIII of the periodic table or a compound thereof with one of Group II and one of Group VI.

IV HOT CARBON REDUCTION:

4 Lepsoe, R., and Kirkpatrick, W. S. Trans. Can. Inst. Mining Met. 40 (in Can Mining Met. Bull. No. 304) 3990404 (1937).
Plants of the Consolidated Mining and Smelting Co. of Canada, Ltd. at Trail, B. C., are described. To produce sulfur, SO_2 and pure oxygen are blown into a water-gas producer.

6 Pavlov, KF., and Andeeva, A. V. J. Chem. Ind. (U.S.S.R.) 14, 1231-2 (1937).
Waste gases containing about 8% SO_2 can be reduced with C; COS, CS_2 and H_2S formed in the reaction can be converted to S.

12 Roesner, Gerhard Chem. Fabrik 1937, 101-6.
Step 1: SO_2 reacts with coke above 900° to form S, COS, and CS_2. Step 2: SO_2, COS and CS_2 are reduced by passing over bauxite or grog at 400-700°.

39 Gazaryan, L. M. Tsvetnye Metal. 1940, No. 8, 69-75.
Flue gases from nonferrous metallurgy operations are treated with hot coke to recover S.

* 40 Lepsoe, Robert Ind. Eng. Chem. 32, 910-18 (1940).
Rates of reduction of SO_2 by C and by CO were studied. To maintain high reduction rates, temperature of the hot zone would not be less than 1300°.

59 Dean, R. S., and Swain, R. E. U. S. Bur. Mines, Bulletin 453, 304 pp. (1944).
The details of operation of the Trail smelter as they affect SO_2 recovery are described. SO_2 and O are blown into incandescent coke to form S, CO_2 and COS.

75 Potts, H. R., and Lawford, E. G. Bull. Inst. Mining Met. No. 509, 1-36 (1949).
S is recovered from smelter gases by the Orkla process at Rio Tinto. Reduction with coke takes place at over 1200°.

112 Amiroya, S. A. Uch. Zap. Permsk. Univ. 13, No. 3, 100-17 (1959).
A thermodynamic analysis of reduction in a coke gas generator is made.

123 Hirose, Kazutomo, et al. Kogyo Kagaku Zasshi 63, 906-11 (1960).
Reduction with coke in a fluidized bed was studied. The effects of temperature, period of contact, flow rate, and the concentrations of SO_2 and O_2 in the starting gas mixtures were examined.

139 Rian, Zun-On. Chosun Kwahakwon Tongbo No. 2, 44-5 (1963). SO_2 in a waste gas was concentrated and then passed over coke at 1100-1300°.

140 Ritter, L. G., et al. Khim. Prom. 1962 (11) 844-7.
141 S was obtained in yields greated than 80% by reduction of SO_2 in gases containing 9-10% SO_2 in a fluidized bed at 950° using coke with a particle size less than 1 mm.

163 Henrich, G. Staub 25, (10) 429 (1965). Methods for flue gas desulfurization are described and compared. Some processes tested in pilot plants and on industrial scale are explained in detail.

293 Lamoreaux, W. F. Eng. Mining J. 113, 198-207 (1922). ZnO was recovered from metallurgical smoke by a Feld scrubber. The cleaned dry gas was passed thru coke at 1100-1300°; recovery of S from the exit gas is discussed.

294 Rassow, B., and Hoffmann, K. J. prakt. Chem. 104, 207-240 (1922) Passage of SO_2 thru glowing charcoal led to the formation of CS_2. The course of the reaction was studied at different temperatures. Above 1100° S and CO are the products of the reaction.

300 Yushkevich, N. F., and Karzhavin, V. A. J. Chem. Ind. (Russia) 2, 719-26 (1926). Theoretical studies of the reduction of SO_2 by C were checked by experiments using coke and birch charcoal. At 800° and above charcoal reduces SO_2 completely no matter at which speed the latter is passed. At 900° coal reduces SO_2 completely provided the gas is passed very slowly.

384 Caddick, A. J. Mining Mag. 53, 213-16 (1935)
COS and CS_2 were also formed when H_2S and elemental S were produced by reducing SO_2. They yielded S in elemental form when mixed with an excess of SO_2 and passed through a tube containing red-hot pumice.

411 Vivian, J. H. Proceedings of the subscribers to the fund for obviating the inconvenience from the smoke produced by smelting copper ores. London, 1833.

415 Fr. Reich. Jahrbuch fur den Berg-und Huttenmann, 1867 p. 158
At a zinc smelter the sulfur dioxide from the blende roasters was decomposed into sulfur by leading it over glowing coal.

431 Scheurer-Kestner. Compt. rend., 114, 296-298 (1892).
In the reaction between sulphate, silica and carbon at a high temperature, the sulphate is converted into a silicate and SO_3. The latter is decomposed by heat into O_2 and SO_2 and these in presence of carbon yield CO_2, CO, and S.

478 Fischer, F. and Pranschke, A. Brennstoff-Chem. 9, 361-3 (1928).
Cokes treated at 300° to 500° in a stream of SO_2 yielded a product containing up to 37.5% S which could not be exacted by solvents.

530 Jolf, H. PB Report 66511 (Frames 173-184 of PB Report 25621). U. S. Dept. of Commerce, Office of Technical Services. May 10, 1937. 12 pages.

541 Kirpatrick, S. D. Chem. Met. Eng. 45, #9, 483-5 (Sept. 1938).
Roaster gas SO_2 was abosrbed in ammonium sulphite solution, removed by treatment with sulfuric acid and reduced to sulfur by means of incandescent coke.

569 Anonymous. Can. Chem. & Proc. Ind. 26, 138-139 (1942).
About 40% of zinc roaster gases was sent to an absorption unit for conversion to ammonium sulphite and then ammonium bisulphite. The latter was treated with sulphuric acid to release SO_2 which was reduced by passage through incandescent coke.

* 575 Siller, C. W., Ind. Eng. Chem. 40, pp 1227-33 (1948).
CS_2 is formed on passing SO_2 over anthracite at 650-1200°, yield 70-90%. Computed free energies of likely reactions are given.

608 Thieler, E. Sulfur. Dresden, Steinkopff, 1936. 132 pages. (Technische Fortschrittsberichte. Volume 38)
This book was cited as containing some information on the reduction of SO_2 to S by carbon.

PATENTS

* 16 Lindblad, Axel R. U. S. Patents 2,090,797 (Aug. 24, 1937)
SO_2 and free oxygen are passed over a carbonaceous material in the presence of an alkali metal compound; in another chamber the oxy-carbon gases formed in step 1 reduce SO_2 under the catalytic influence of condensed alkali metal compound.

17 Lepsoe, Robert (to Consolidated Mining & Smelting Co. of Canada, Ltd.) U.S. 2,102,081 (Dec. 14, 1937)
The SO_2 is reduced by a fuel bed of carbonaceous material; the temperature of the reaction is controlled by admitting at high levels of the fuel bed additional SO_2 at a temperature below the prevailing temperature of the reaction gases.

19 Pourbaix, Marcel, and Societe generale industrielle at chimique du Katanga (Sogechim). Brit. 471,668 (Sept. 2, 1937)
The reduction is effected with C at 800° in such a way that in the resultant gas the ratio $SO_2/(COS+H_2S+CS_2)=$ substantially 1.

27 Von Giresewald, Conway, and others (to American Lurgi Corp.) Can. 384,396 (Oct. 3, 1939).
Reduction is effected by a layer of fuel in a shaft furnace. The effluent is treated with SO_2 for recovery of additional S and conversion of CO to CO_2.

35 I.G. Farbenind. A.-G. (Hans Dohse and Herbert Wolf, inventors) Ger. 670,884 (Jan. 26, 1939)
Reduction is effected by coke at 700° in a shaft furnace. The resulting gas is subjected to several treatments to recover up to 90% S.

* 48 Carter, Bernard M. (to General Chemical Co.) U. S. 2,208,247 (July 16, 1940)
A multi-stage reduction process using a solid carbonaceous reducing agent.

50 Von Girsewald, Conway Frhr. and others (to Metallgesellschaft A.-G.) Ger. 711,509 (Sept. 4, 1941)
After reduction with a carbonaceous material at an elevated temperature, the gas stream is treated with SO_2 and oxygen at 400-800° in the presence of a catalyst. The resulting gases are treated with another catalyst.

58 Byrns, A. C. (to Union Oil Co. of California) Can. 413,254 (June 15, 1943)
The catalyst is Co molybdate or thiomolybdate upon a porous gel of hydrous Al, Zr, Ti or Th oxide.

69 Ferrante, Corrado. Ital 412,812 (Feb. 16, 1946),
Sulfurous gases pass through a red hot layer of coal where S is formed, but it reacts with CO to form COS and CS_2. In the next stage the gas mixture reacts with other SO_2, using a catalyst such as bauxite, Fe_2O_3 and $CaSO_4$.

88 Mukoyama, Mikio, and Futakuchi, Magoichi. Japan. 5174
October 9, 1953. A mixture of SO_2 and steam is pass thru powdered charcoal at 1050°.

92 Sakamoto, Takeshi (to Tohoku Fertilizer Co.) Japan. 6866
October 23, 1954
After being concentrated by means of dimethylaniline, SO_2 is passed into a reducing furnace where it is mixed with air and coke; reduction to S, CS_2, H_2S, CO, CO_2 and N takes place at 1000-1200°.

248 Carpenter, Franklin R. U.S. 925,751 (June 22, 1909).
Gases are passed over glowing C, further heated to reduce sulfides, and then cooled to condense S.

258 Ruthenburg, M. Brit. 29,030 (Dec. 17, 1912)
SO_2 is reduced in a column of carbonaceous material heated electrically.

260 Thiogen Co. Ger. 271,002 (Nov. 24, 1912).
Vapors containing SO_2 are allowed to act on hot C in the presence of metal compounds.

264 Basset, L. P. Brit. 20,716 (Sept. 13, 1913)
Gases are brought in contact with incandescent C, whereby a mixture of CS_2, COS and CO is produced; this mixture is caused to react with excess of SO_2 in a 2nd reaction chamber.

265 Young, S. W. Can. 147,882 (May 13, 1913)
Gases are subjected to the action of a heated carbonaceous fuel in presence of a solid reagent capable of neutralizing the SO_2.

271 Lamoreaux, W. F., and Renwick. C. W. U. S. 1,140,310 (May 18, 1915).
Smelter gases are passed through an electrically heated bed of incandescent coke.

273 Lamoreaux, W. F., and Renwick, C. W. Brit. 2,834 (Feb. 22, 1915)
Furnace gases are passed thru coke heated to over 1000° by means of an electric current.

274 Lamoreaux, W. F. Brit. 16,543 (Nov. 23, 1915)
Heated gases are passed over incandescent coke. Apparatus is described.

278 Lamoreaux, W. F. U. S. 1.182,915 (May 16, 1916).
Smelter gases are mixed with CO, H_2S, H or hydrocarbon vapor, and the mixture is passed over a bed or incandescent coke which is heated electrically.

282 Anderson, A. G. Swed. 42,657 (June 13, 1917)
Gases from roasting furnaces are divided into two portions; one is passed over glowing coal, the other is passed thru a regenerator where heat is absorbed. Then the two portions are mixed in the regenerator, partial reduction to S taking place.

291 American Smelting & Refining Co. Brit. 144,306 (June 3, 1920)
Sulfur dioxide, separated from metallurgical gases, is mixed with air; the mixture is heated and then passed over incandescent coke from where it goes to a settling chamber.

295 Howard, G. C. U. S. 1,417,068 (May 23, 1922); Can. 219,617 (May 13, 1922).
Liquefied SO_2 is vaporized, mixed with air, and the mixture passed over incandescent coke.

306 Howard, W. H. U. S. 1,685,231 (Sept. 25, 1928)
A current of air is utilized to take from aqueous solution the SO_2 absorbed from furnace gases and the mixture of air and SO_2 subjected to the action of incandescent carbonaceous material to form S.

* 310 **Thompson, A. P.** (to General Chemical Company) U. S. 1,695,068 (Dec. 11, 1928).
In reducing SO_2 by the action of carbonaceous fuel, an endothermic reaction such as that of CO_2 mixed with SO_2 is simultaneously effected in the absence of H in any form, so that the temperature of the reduction process is controlled.

311 Patentaktiebolaget Grondal-Ramen. Fr. 659,095 (Aug. 17, 1928).
Ores such as pyrites are fused in a blast furnace having its mouth closed, with reducing material such as coke, so that SO_2 formed is reduced to S and O is transformed into CO or CO_2.

320 Bacon, R. F. U. S. 1,734,991 (Nov. 12, 1929).
Roaster gas is compressed and subjected to the action of hot reducing gas such as that from glowing coal or coke.

* 321 Benner, R. C. and Thompson, A. P. (to General Chemical Co.) U. S. 1,771,480 (July 29, 1930).
The gaseous products, formed by heating a SO_2 containing gas with bituminous coal, are passed through a bed of incandescent coke and the resulting gaseous products are subsequently passed over a catalyst such as French bauxite to complete the reduction of SO_2 to S.

322 Benner, R. C. (to General Chemical Co.) U. S. 1,773,294 (Aug. 19, 1930).
SO_2 is reduced with carbonaceous fuel to produce a gas mixture containing elemental S, and the composition of the gas mixture is regulated to provide approx. interacting proportions of reducing and reducible gases and substantial quantities of CO_2 are preferentially oxidized.

323 Tyrer, D. and Imperial Chemical Industries, Ltd. Brit. 350,124 (April 5, 1930).
Heated SO_2 containing gases are passed into contact with heated coke at a high temperature with the formation of CO and CO_2.

325 Clark, A. M. and Imperial Chemical Industries, Ltd. Brit. 355,832. (May 31, 1930).
Elemental S is obtained and separated from gases such as those resulting from the reduction of SO_2 by coke at a high temperature, by subjecting the mist produced by cooling the gases to an aging period at low temperatures.

326 Tyrer, D. and Imperial Chemical Industries, Ltd. Brit. 357,178 (June 16, 1930).
Low-temperature coke (semi-coke) is used for reducing SO_2 to S and reaction temperatures of 600-1000° are suitable.

327 Tyrer, D. and Imperial Chemical Industries, Ltd. Brit. 358,580 (July 11, 1930).
The gases are passed through a bed of coal at carbonizing temperature (at least 400-500°) and the gases and vapors thus produced are then passed through a bed of highly heated coke, produced by carbonization of the coal, with sufficient air to maintain a temperature of 1000-1200° to complete the reduction of SO_2 and crack tarry materials.

338 Benner, R. C. and Thompson, A. P. (to General Chemical Co.) U. S. 1,836,357 Dec. 15, 1931.
Proportions of SO_2 and O in entering and added gas mixtures are adjusted to provide approx. reacting proportions of reducing and reducible gases and to have the major portion of the C utilized for the production of CO.

339 Bacon, R. F. Brit. 364,351, Jan. 12, 1931
Recovery of elemental S from roaster gases is facilitated by preheating, preferably to 300-600°, the air used for treating the sulfide ore, which increases the SO_2 in the gases and renders the reduction by coal or coke more efficient.

347 Baum, K. M. (to United Verde Copper Co.). Brit. 369,956. (March 29, 1932).
In the formation of S by reaction between H_2S and SO_2 the catalyst is a mixture of adsorbent materials, preferably activated C and a mineral gel, e.g., silica gel or a zeolite, in the ratio 2:1.

352 Lenander, N. E. (to Texas Gulf Sulphur Co.). U. S. 1,904,481, (April 18, 1933).
Pyrites is smelted with coke and flux in a closed shaft furnace provided with means for introduction of an oxidizing gas for producing molten slag and a gaseous product containing elemental S vapor, SO_2 and other S compounds; the gaseous product is withdrawn from the furnace and is passed at a temperature of about 400° through a chamber containing a catalyst such as bauxite or suitable metal oxides. U.S. 1,904,482 relates to a process in which S in produced by

the reaction of SO_2 and CS_2 gases at an elevated temperature in the presence of a catalyst such as bauxite. U.S. 1,904,483 relates to S production by reaction between a gaseous S compound such as H_2S, CS_2 or COS and another gas such as SO_2 at an elevated temperature in the presence of a catalyst comprising Al hydrate embodied in a porous cement-like product.

354 Tyrer, D. (to Imperial Chemical Industries, Ltd.).
U. S. 1,908,873, (May 16, 1933).
A mixture of SO_2 and air is passed through a bed of hot semi-coke and the resulting hot reaction gases are passed over raw coal to carbonize it to semi-coke and the resulting mixture of gases is passed over hot coke obtained in the first-mentioned step of the process.

357 Bacon, R. F. and Hotchkiss, H. T., Jr. (to R. F. Bacon)
U. S. 1,917,789, (July 11, 1933).

359 Bacon, R. F. U. S. 1,934,474, (Nov. 7, 1933).
Sulfide ores are roasted in the presence of regulated amounts of air and under pressure and the resulting SO_2-bearing gas, without any substantial loss of heat, is subjected to reduction under super-atmospheric pressure in the presence of incandescent carbon.

* 360 Tyrer, D. (to Imperial Chemical Industries Ltd.) U.S. 1,936,809 (Nov. 28, 1933).
Oxygen is added to the SO_2-containing reaction gases from a low temperature carbonization of coal and the mixture reduced at high temperature (1000°) by passage over carbonaceous fuel.

362 Bacon, R. F. Brit. 389,520, (March 20, 1933).
In the blast furnace smelting of pyrites or other sulfide ores for the production of S, insufficient coke being used to reduce the SO_2 produced, the SO_2 is brought into contact with sufficient hot sulfide to reduce it to S within the furnace. The reaction is facilitated by the addition of CaO or MgO.

363 Baum, K. M. (to the United Verde Copper Co.). Can. 322,208, (May 2, 1933).
Sulphur is recovered by subjecting hydrogen sulphide to the action of sulfur dioxide in the presence of particles of activated carbon impregnated with a mineral gel.

374 Levy, S. I. Brit. 403, 961,(Feb. 26, 1934.)
Pyrites ore is roasted with so much air that only part of the S is oxidized. The temperature of the outgoing gases exceeds 1000°. Reduction of the SO_2 is effected in another chamber by finely divided coal, coke or charcoal so as to form CO_2 and S.

375 Tyrer, D. and Imperial Chemical Industries, Ltd. Brit. 406,343, (Feb. 26, 1934).
Roaster gases are treated for recovery of S by reaction with a solid, liquid or gaseous carbonaceous reducing agent, e.g., glowing coke. The resultant gases are passed with an excess of SO_2 over a catalyst to convert H_2S, COS and CS_2, formed during the reduction, to S.

* 376 McCluskey, S. B. Brit. 417,276,(Oct. 2, 1934.)
The process of reducing the SO_2 in roaster gases is made cyclic by controlling the concentration of SO_2 and incompletely reducing it.

377 McCluskey, S. B. Brit. 421,290, (Dec. 18, 1934).
In the reduction of SO_2 of gaseous mixtures, the gas to be treated is passed through a comparatively shallow layer of solid carbonaceous fuel disposed on a chain or other mechanically actuated grate.

380 Imperial Chemical Industries, Ltd. Fr. 700,064,(Feb. 16, 1934).
Gases containing SO_2 are reduced by a carbonaceous reducing agent and the resultant gases are passed, with an excess of SO_2, over a catalyst to convert the COS and H_2S into sulfur.

386 Benner, R. C. and Thompson, A. P. (to General Chemical Co.)
U.S. reissue 19,531, (April 16, 1935).
A reissue of original patent No. 1,771,480.

* 390 Lindblad, A. R. Brit. 426,456, (April 3, 1935).
For the reduction of SO_2 over heated carbonaceous fuel, the composition and speed of passage of the preheated gas mixture was controlled so that the carbonaceous gas was chiefly CO. A rotary grate was used.

393 Lindblad, A. R. Brit. 428,979, (May 22, 1935).
Roaster gases are cooled and passed through an electrostatic precipitator before high temperature reduction (preferably with catalyst) over coal or coke.

394 Lindblad, A. R. Brit. 437,236, (Oct. 25, 1935).
The gas mixture obtained by passing roaster gases over carbonaceous material and an alkali metal compound is mixed with a regulated amount of SO_2 and led to a chamber filled with catalyst, e.g., brick or bauxite.

395 Siemens-Lurgi-Cottrell-Elektrofilter-Gesellschaft fur Forschung und Patentverwertung. Brit. 439,467, (Dec. 6, 1935).
SO_2-containing gases leaving the reduction stage are first ionized and the impurities are then precipitated in a separate electric field constituted by two nonionizing electrodes, one of which is constructed as a catalyst surface.

396 Carter, B. M. (to General Chemical Co.). Can. 352,653, (Aug. 27, 1935).
SO_2 is reduced to S by contact at 425-650° with carbonaceous residues formed by the destructive distillation of acid sludge.

397 International Mining Trust Registered (Serge Diakonoff, inventor). Fr. 775,910, (Jan. 12, 1935).
SO_2 from roaster gases is reduced with C in the presence of steam and causing the H_2S to react with SO_2 to form S and H_2O. The SO_2 is concentrated by absorption and reevaporation.

398 Metallgesellschaft A.-G. Fr. 784,942, (July 29, 1935).
SO_2 is withdrawn from gases by a mixture of organic bases, such as xylidine, with water, in the presence of substances capable of decomposing the difficulty soluble sulfates of the organic bases. The SO_2 is driven off again by heating, caused to react at 800-1200° with a carbonaceous reducing agent, submitted to a complementary treatment at 400-700° by catalysts to destroy attached S compounds, and the S is precipitated by a high-tension electrostatic field.

* 405 Lindblad, A. R. Brit. 443,008, (Feb. 19, 1936).
The SO_2 in a gas mixture which contains O also is reduced in a multi-stage process.

408 Siemens-Lurgi-Cottrell Elektrofiltergesellschaft m.b.H. fur Forschung und Patentverwertung. Fr. 796,155, (March 31, 1936).
SO_2 is reduced by means of C, passed through an electric filter to remove impurities and then over a catalyst such as alkalies, Al or Fe.

409 Patentaktiebolaget Grondal-Ramen. Ger. 628,554, (April 8, 1936). S, or As_2O_3 and Cu, obtained by smelting pyrites in a furnace with insufficient carboniferous reducing agent (coke) to reduce the SO_2 and As_2O_3 formed, escape with the gases to be reduced later by gases containing CO.

413 De Sincay, A.C.S. Brit. 2993,(Nov. 21, 1865). Roaster gases are reduced by passage over coal, coke or other carbon material.

414 De Sincay, A.C.S. U.S. 59,328, (Oct. 30, 1866). Same as #413.

416 Hartmann, W. E. A. Ger. 9275 (1880). H_2S was produced by the reduction of SO_2.

429 Hanisch, E. and Schroeder, M. U.S. 361,761, (April 26, 1887). Roaster gases are passed through water. The SO_2 is driven from solution by heating and reduced over a glowing bed of fuel.

455 Vadner, C. S. U.S. 1,103,165, (July 4, 1914). Smelter fumes are mechanically purified, then mixed with hydrocarbon gases which have been passed through a column of red-hot coke, and the mixture heated to a high temperature in a gas-converting chamber. The products are cooled to remove moisture and part of the impurities, and then passed through scrubbers of lime rock and ferric oxide into receptacles for use for power and fuel purposes.

456 Vadner, C. S. Ger. 276,368, (July 15, 1914). Same as #455.

460 Lamoraux, W. F. and Renwick, C. W. Fr. 477,793, (Feb. 25, 1915). SO_2 is reduced by contact with incandescent carbon and the required heat is furnished independently of the heat of combustion of C.

462 Lamouraux, W. F. Fr. 480,345, (July 19, 1916). SO_2 is reduced by passage over incandescent carbon.

467 Dow, H. H. and Strobacker, C. J. U.S. 1,350,858, (Aug. 24,1920). CS_2 is made by bringing SO_2 into contact with heated coke and adding sufficient oxygen with the SO_2, to maintain the coke at the required temperature.

476 Riedel, F. Ger. 425,664 (Feb. 23, 1926).
Gases containing sulfur compounds, alone or mixed with air, are used in place of air in coal dust or oil furnaces.

479 Howard, W. H. U.S. 1,685,231 (Sept. 25, 1928)
Same as #306.

487 Tyrer, D. and Imperial Chemical Industries, Ltd. Brit. 358,580, (July 11, 1930).
SO_2 is reduced in a two-stage procedure.

499 Tyrer, D. (to Imperial Chemical Industries, Ltd.). U.S. 1,908,873, (May 16, 1933).
Same as #354.

* 509 Gleason, G. H. and Loonam, A. C. (to Guggenheim Bros.) U.S. 1,972,883, (Sept. 11, 1934).
A gas mixture such as a smelter gas in contacted with a SO_2 absorbing medium, such as a soln. of NH_4 sulfite and bisulfite, the SO_2 is subsequently released and reduced by passage over hot carbon. U.S. 1,972,884 relates to a process of S condensation in which the S vapor is passed through molten S maintained below 150°. U.S. 1,972,885 relates to a process involving the treatment of a gaseous mixture containing S vapor to remove deleterious contaminants.

511 McClusky, S. B. Ind. 20,939, (Nov. 10, 1934).
Same as #377.

512 Lindblad, A. R. Swed, 80,439, (May 23, 1934).
SO_2 from pyritic materials is reduced by means of coal or other solid carbonaceous material impregnated with or mixed with alkali metal compounds.

517 Bolideas Gruvaktiebolag (P.H.W.Agren and H.H.G.Lindhe, inventors). Swed. 84,978, (Dec. 3, 1935).
SO_2 us reduced by means of carbon in presence of a catalyst prepared from a mixture of TiO_2 or other Ti compound and cement, e.g. s0-called aluminate cement.

520 Bacon, R. F. and Judson, W. U.S. 2,040,682, (May 12, 1936).
Finely divided material such as pyrites together with finely divided coke are passed in suspension through a countercurrently flowing gaseous product containing S vapor at a temperature above 800°, to liberate the S of the metal sulfide

and produce a final gaseous product containing a substantial proportion of recoverable S vapor, and the residual material and coke in suspension are passed through a countercurrently flowing gaseous product containing SO_2 at a temperature above 800°, so that the SO_2 is almost completely reduced by the coke and the gaseous product containing S vapor mentioned is produced, and the residual material in suspension is then passed through a countercurrently flowing oxidizing gas at a temperature above 950° to produce oxide from the remaining metal sulfide and to produce t[illegible]as mentioned containing SO_2.

525 Lepsoe, R. and Mills, J. R. (to Consolidated Mining and Smelting Co. of Canada, Ltd.). U. S. 2,080,359, (May 11, 1937). A sulfur dioxide reducing catalyst is produced from a molded and dried mixture containing $Al(OH)_3$, a refractory carrier material such as fire clay, a solid combustible material such as sawdust, and a sintering agent such as Na silicate by sintering the material at a temperature sufficiently low to retain at least a part of the water of hydration of the $Al(OH)_3$. U.S. 2,080,360 (Lepsoe, R. (to consolidated Mining and Smelting Co. of Canada, Ltd.) relates to a process for the production of substantially pure S from SO_2 by introducing the SO_2 into a carbonaceous fuel bed to form a gas mixture containing CO and S, subjecting the gas mixture to such conditions that substantially all the convertible S combines to form C oxysulfide, cooling the gas to below the condensation point of S, separating solid and condensed impurities from the C oxysulfide, adding clean SO_2-bearing gas to convert C oxysulfide to S and CO_2.

526 Carter, B. M (to General Chemical Co.). U.S. 2,080,384, (May 11, 1937).
Same as #396.

562 A. G. Exploration Co. Ltd. Brit. 525,730, (Sept. 3, 1940). Gases containing up to 16% SO_2 less than 1% O are reduced at 1000° or higher by passage through a column or mass of coke or charcoal at such a rate that the carbonaceous material is completely consumed.

564 Lindblad, A. R. Ger. 693,468, (June 13, 1940).
Gases containing SO_2 are passed through a generator charged with powdered coke, charcoal or brown coal, or a mixture of these. The charge is saturated with Na_2CO_3, Na_2SO_4 or Na_2S

and is kept at 700-1100°. From the generator the gases are led through a reaction chamber where the S separates. If gases contain traces of water vapor CO, COS, CS_2 and H_2S will form.

565 Pyrites Co., Ltd. Ital. 380,818 (Feb. 22, 1940).
Gases containing not less than 12% SO_2 are reduced by passage over coke at more than 850% in a closed furnace.

567 Metallgesellschaft A.-G. Ger. 715,845, (Dec. 4, 1941).
SO_2 is almost completely reduced by passage through a bed of hot coal in a furnace. To the gases leaving the coal bed, at less than 800°, is added the required quantity of SO_2 to oxidize CO to CO_2 and convert COS to CO_2 and S. The gaseous mixture is passed over suitable catalysts at a temperature above the m.p. of S, up to approx. 700°.

568 Metallgesellschaft A.-G. Nor. 63,291, (March 3, 1941).
The gases are dried by means of H_2SO_4 or other drying agent, then reduced at high temperature with C-containing materials and finally after-treated with SO_2 in the presence of catalysts at about 400-800°.

594 Lamoreaux, W. F. U. S. 1,169,726, (Jan. 25, 1916)..
Same as #460.

596 A.-G. fur Anilinfabrikation. Ger. 300,715, (Jan. 2, 1920)
The volatile sulfur compounds present after reduction of SO_2 over carbon in a blast furnace are caused to react with calcium sulfate with the formation of carbon dioxide, water, sulfur and calcium sulfide. Recovery of the bound sulfur is facilitated by addition of iron oxide, alumina, clay granite, etc.

597 MacLeod, J. G. from Browning, W. J. Brit. 158,288, (Feb. 10, 1921.)
Gases from pyrites combustion are passed through incandescent carbonaceous matter, the heat of which is maintained by its own combustion. A regulated quantity of oxygen and, if necessary, water vapor is mixed with the gases.

598 Browning, W. J. U.S. 1,407,323, (Feb. 21, 1922).
Hydrogen sulfide is obtained by inter-reaction of added hydrogen with the sulfur obtained by reduction over incandescent carbon of the SO_2 in sulfurous gases to which air and hydrogen have been added.

607 Siemens-Lurgi-Cottrell-Elektrofilter-Gesellschaft fur Forschung und Patentverwertung. Brit. 439,467, (Dec. 6, 1935).
Same as #395.

610 Tharaldsen, F. Nor. 55,862, (June 29, 1936).
The SO_2 together with sulfur vapor is led through a layer of glowing coal.

612 Lindblad, A. R. Can. 363,638, (Jan. 26, 1937).
A gas mixture containing SO_2 and Oxygen, obtained by roasting pyrites is first reduced in a gas generator by carbon which has been mixed with alkali compounds, then, in the hot state, led together with a regulated amount of extra SO_2 through a reaction chamber containing a material which acts catalytically.

613 Lepsoe, R and Mills, J. R. (to Consolidated Mining and Smelting Co. of Canada, Ltd.). Can. 367,891, (Aug. 10, 1937)
Same as #525, part I - U.S. 2,080,359

627 Metallgesellschaft A.-G. Belg. 520,290, (June 10, 1955).
In the preparation of sulfur by catalytic reaction of hydrogen sulfide with oxygen or sulfur dioxide at least two catalysts are used. One is held above the dew point of S and the second below the dew point of S. S deposits on the two catalysts and is removed by heating before the catalyst becomes inactive.

644 Akbar, M. (Shell Internationale Research Maatschappij N. V.). Brit. 1,138,437, (Jan. 1, 1969).
A hydrogen sulfide-containing feed gas is burned with air or another O-containing gas to produce a hot gas containing S, SO_2 and H_2S. The gas is cooled and S is condensed and removed. The remaining gas is heated for the subsequent catalytic reaction between H_2S and SO_2.

V HYDROCARBON REDUCTION:

24 Zawadski, J., Ostrouch, S. and Kwiecinski, G. Przemysl Chem. 22, 558-64 (1938).
The dependence of the course of the reaction between CH_4 and SO_2 (taking place in a quartz tube, filled with fused quartz pieces) upon the temperature and the composition of the initial mixture was investigated.

* 114 Baccarreda, M. et al. Chem. e ind. (Milan) 41 pp. 277-80 (1959).
CH_4/SO_2 in 1.3 to 1.5 ratio react on kaolin catalyst to yield CS_2 at 700-800°. One pass yield of 54-57% is reached. Normally 1/2 mol H_2S and 1 mol S are formed per mol CS_2.

* 67 Walker, S. W. Ind. Eng. Chem. 36, 906-12 (1946).
Equilibrium calculations from existing data give a clear picture of the effect of certain variables on the reaction of SO_2 and CH_4. Experimental confirmation of the conditions predicted by the calculations is reasonably satisfactory.

* 76 Fleming, E. P. and Fitt, T. C. Ind. Eng. Chem. 42, 2249-53 (1950).
A description of the method for the high temperature combustion of approx. 7% SO_2 gas with an automatically controlled volume of natural gas.

* 108 Kulcsar, J. G. et al Studia Univ. Nictor Babes et Bolyai 3, No. 4, Ser. 1, No. 2, 207-21 (1958).
The reduction of sulfur dioxide with methane using a silica gel catalyst was studied.

* 154 Vilesoy, N. G. et al. Tr. Mosk. Khim-Tekhnol. Inst. No. 47, 86-9 (1964) (Russ.)
CS_2 was produced at a 50% yield by reacting sulfur dioxide and methane in the presence of lead phosphate deposited on chamotte brick.

* 195 Vilesoy, N. G. and Gorbatykh, G. A. Khim. Prom. 42 (3), 187-9 (1966) (Russ.)
Equilibrium compositions of the gas mixtures obtained in the sulfur dioxide reduction by methane were calculated at 427-670°C.

334 Young, G. J. Eng. Mining J. 131, J (1931).
Sulfur produced by reduction of SO_2 by CH_4 or CO in an electrically heated laboratory reactor in the presence of a simple catalyst.

* 369 Yuskevich, F. et al. J. Chem. Ind. (Moscow) 1934, No. 2, 33-7.
CH_4 and SO_2 reacted at 900°C over a bauxite catalyst to form S, H_2S, H_2O and CO_2. Benzine also reacted with SO_2 at 700-800°C to produce sulfur at 60-90% yields.

465 Papish, J. Proc. Indiana Acad. Sci. 1918, 170-1.
Experiments show that SO_2 is reduced to sulfur by methane of elevated temperatures. Sulfur occurring in volcanic regions may be formed in this manner.

* 222 Shakhtakhtinskii, G. B. et al. Azerb Neft Khoz 46 pp 40-2 (1967 Russia).
A comparison of bauxite and alunite catalysts in nat. gas (95% CH_4) reduction of SO_2 in terms of sulfur yield (82-83% single stage, up to 98-99% two stage).

* 616 Cole, C. S. Reduction of sulfur dioxide with hydrocarbon gases. M. S. Thesis, University of Illinois, 1940.
The reductants and catalysts for SO_2 reduction were discussed. SO_2 was experimentally reduced to sulfur by natural gas over a calcium sulfate catalyst.

PATENTS

15 Merriam, H. F. (to General Chemical Co.). U. S. 2,090,217. (Aug. 17, 1937).
Apparatus and operative details described for reduction of SO_2 by natural gas in a reactor wherein mounds of catalytic material are spaced apart in the checker work.

* 53 Fleming, E. P. and Fitt, T. C. (to American Smelting and Refining Co.). U. S. 2,270,427, (Jan. 20, 1942).
The equipment used for reduction of SO_2 with natural gas is detailed. This consists of a high temperature reactor, an intermediate converter and two claus converters and the requisite heat exchangers and sulfur condensers.

* 65 Fleming, E. P. and Fitt, T. C. (to American Smelting and Refining Co.). U. S. 2,388,259, (Nov. 6, 1945).
This describes the conditions used in the process equipment described in #53, i.e. high temperature reduction of 1200°C, intermediate conversion at 425-450°C, low temperature reaction at 200-250°C and sulfur condensation at 120-140°C.

* 71 Fleming, E. P. and Fitt, T. C. (to American Smelting and Refining Co.). U. S. 2,431,236, Nov. 18, 1947.
Modification of #65.

* 100 Shiromoto, Y. and Hoshiba, Y. (to Fuji Iron & Steel Co.). Japan, 6615 (Aug. 7, 1956)
Active H produced by thermal decomposition of hydrocarbon gases reduces SO_2 to H_2S at 1200°C. H_2S and SO_2 can be further reacted to produce sulfur.

* 118 Metallgesellschaft Akt.-Ges. (by W. Thumm). Ger. 1,051,258, (Feb. 26, 1950).
Hydrocarbons and oxygen are utilized to provide the heat necessary to produce CO and H and partial reaction of SO_2 at $>$ 1200°C. The CO, H, and remaining SO_2 react to produce sulfur at 900-1000°C.

* 176 West, J. R. and Conroy, E. H. (to Texas Gulf Sulphur Co.). U. S. 3,199,955, (Aug. 10, 1965).
Describes a process similar to #65 except that the high temperature reduction takes place below 1000°C in the presence of a catalyst such as activated alumina, bauxite, calcium sulfide or quartz.

* 238 Aarvik, J. (to Norges Geologiske Undersokelse). U. S. 3,369,872, (Feb. 10, 1968).
SO_2 is reduced to sulfur by mineral oil using cordierite or zeolites as catalysts.

254 Teed, F. L. et al. Brit. 14,628, (June 20, 1911).
SO_2 is reduced at 750-1000°C with reducing gases such as H, hydrocarbons, CO, coal gas, water gas, or producer gas with catalysts; i.e. spongy Pt, platinized pumice, MgO, $CaSO_4$, Iron oxides, etc.

358 Merriam, H. F. (to General Chemical Co.). U. S. 1,919,342, (July 25, 1933).
Details of apparatus arrangement and operation are described for absorption of SO_2 from burner gas in water and the subsequent reduction of the absorbed SO_2 to sulfur using natural gas and a bauxite catalyst.

379 Rosenstein, L. (to the United Verde Copper Co.). Can. 342,308, (June 12, 1934)
Hydrocarbons are employed to reduce SO_2 in gaseous mixtures to hydrogen sulfide using a catalyst at elevated temperatures. Sulfur is produced by reacting the H_2S with additional SO_2.

403 Merriam, H. F. (to General Chemical Co.) U. S. 2,043,202, (June 2, 1936).
In the reduction of SO_2 with natural gas at elevated temperatures, a portion of the reaction product gases are recirculated to control the reaction temperature.

407 Ahlqvist, H. Can. 356,673, March 24, 1936
The product gases from the reaction of H_2S and SO_2 are cooled by a water spray to condense the sulfur. The H_2S for this reduction is produced by reacting 2/3 of the sulfur product with hydrocarbon oil at high temperatures.

461 Garner, J. B. and Clayton, H. D. U. S. 1,173,366, (Feb. 29, 1916). SO_2 is combined with natural gas or producer gas or other gaseous fuel containing H, in the presence of wood charcoal at 500-600°C to form H_2S.

463 Garner, J. B. and Clayton, H. D. (to Metals Research Co.). U. S. Reissue 14,277, (March 27, 1917), Original No. 1,173,566. Same as #461.

482 Benner, R. C. and Thompson, A. P. (to General Chemical Co.). U. S. 1,741,551, (December 31, 1929). SO_2 is reacted with a reducing fuel at 350-450°C. After condensation of sulfur, further reaction to produce sulfur is obtained by employing an activated bauxite catalyst below 200°C to effect SO_2 reduction with H_2S or COS.

507 Rosenstein, L. (to United Verde Copper Co.). U. S. 1,967,264 (July 24, 1934). SO_2 is reduced by methane to give H_2S by using manganese sulfide as a catalyst at 700-850°C.

515 Merriam, H. F. (to General Chemical Co.). Can. 352,504, (Aug. 20, 1935). Same as #15.

532 Oliver, T. C. (to Chemical Construction Co.). U. S. 2,068,430, (Jan. 19, 1937). Iron sulfide ores are oxidized to produce molten iron oxide and SO_2. The SO_2 is reduced to sulfur, using natural gas. Residual SO_2 in the waste gas is absorbed in a liquid, and then desorbed and used as a portion of the combustion gas for the initial step.

629 Viles, P. S. (to Esso Research and Engineering Co.). U. S. 2,887,363 (May 19, 1959). H_2S is produced when SO_2 and an excess of methane are reacted at 1100-1300°C in the presence of cobalt molybdate catalyst (fixed-bed or fluidized) supported on alpha or gamma alumina.

* 645 Van Helden, H. J. A. (Shell Internationale Research Maatschappij N. V. S. African 68 01,298 (July 18, 1968). SO_2 is absorbed from an SO_2-O_2 mixture by a copper oxide acceptor, the SO_2 is then reduced by methane or propane at 350-450°C over a catalyst prepared by impregnating bauxite, alumina or silica-alumina with vandaium.

VI H_2S - NORMAL CLAUS:

41 Udintseva, V. S. and Chufarov, G. I. J. Chem. Ind. (U.S.S.R.). 17, No. 3, 24-8 (1940). The effect of the wall surface (glass, alumina, iron, iron oxide or copper) on the non-catalytic reaction.

78 Murthy, A. R. V. and Rao, B. S. Proc. Indian Acad. Sci. 34A, 283-98 (1951).
Kinetics of the gaseous reaction between hydrogen sulfide and sulfur dioxide was studied using Ag_2S-CO thiomolybdate, COS, and Mo sulfide as catalysts, the partial pressure of H_2O vapor was maintained constant.

134 Averbukh, T. D. et al Khim. Prom. 1962, No. 4, 51-8.
Desulfurization of waste gases in copper smelters with sulfur recovery applying additional combustion; additional circulation of waste gas containing 3% CS_2, COS, H_2S and SO_2 through catalyst in special furnaces reduces air pollution and produces sulfur.

138 Perepelkin, K. E. et al. Khim. Volokna 4 (2), 38-40 (1962).
The flue gases from the manufacture of carbon disulfide are oxidized in two stages to produce sulfur. In the first stage, reaction occurs without a catalyst at 700-1000°C. The Claus reaction is used for the second stage.

175 Winning, M. D. Can. Mining & Met. Bull. 58 (642), 1097-1102 (1965).
A review of the sulfur recovery industry in western Canada. A brief technical description of a typical modified Claus sulfur recovery unit is included.

181 Chute, A. E. European Chem. News Large Plant Supp. (London) 21-3 (Sept. 30, 1965).
The demand for sulfur has resulted in the development of large-scale catalytic units for the recovery of sulfur from hydrogen sulfide. New large-scale plants are listed, and production costs analyzed.

191 Opekar, P. C. and Goar, B. G. Hydrocarbon Process. Petrol. Refiner 45 (6), 181-5 (1966).
A computer program optimizes the design and operation of a Claus-type sulfur plant.

203 Chute, A. E. Petr/Chem. Engr. 39 (6), 32-6 (1967).
Describes the necessary steps for using the Claus process to recover sulfur from high sulfur crudes and natural gas and to control air pollution by decreasing the H_2S content of the stack gas to 10 ppm or less.

206 Fischer, H. Chem.-Ing. Tech. 39 (9/10), 515-20 (1967) (Ger.).
Provides information for a wide range of conditions for utilization of the Claus reaction. Includes flow diagrams, processes for varied strength H_2S feed gases, varied plant capacities and some cost and operating expense data.

214 Konopatov. A. P. and Strom, L. D. Neftepererab. Neftekhim. 1967 (2), 25-9 (Russ.). H_2S is extracted from refinery gases by means of a monoethonolamine solution. The H_2S is desorbed by steam stripping. The H_2S is converted to sulfur by partial oxidation of 2/3 of the H_2S to SO_2 and oxidation of the remaining H_2S with the SO_2 thus formed.

216 Munro, A. J. E. and Masdin, E. G. Brit. Chem. Eng. 12, 369-73 (March 1967).
A study of the Claus method for desulfurizing fuel gases using a molecular sieve catalyst.

246 Carpenter, F. R. Proc. Colorado Sci. Soc. 9, 65-78, (1909).
Describes four methods of eliminating SO_2 from a smelter near Salt Lake City, (1) Liquefaction of SO_2, (2) Reduction of SO_2, (3) Conversion of a part of the SO_2 to H_2S (for the Claus reaction), (4) Conversion of all the SO_2 to H_2S.

305 Taylor, H. A. and Wesley, W. A. J. Phys. Chem. 31, 216-30 (1927).
Data for the gaseous reaction of SO_2 and H_2S in Pyrex glass at 370-730°C shows that the rate is proportional to the surface area of the reaction chamber.

317 Tominaga, H. and Toraishi, W. J. Chem. Soc. Japan 50, 699-705 (1929).
The reaction velocity between SO_2 and H_2S has a minimum value between 0° and 100°C. The reaction velocity is increased when the apparatus contains adsorbed water.

343 Yushkevich, N. F. et al. J. Chem. Ind. (Moscow) 1932, No. 3. 17-26.
Equilibrium calculations show that H_2S and SO_2 react to completion below 200°C but the rate is very slow. At 220-250°C the reaction proceeds at a satisfactory rate using a natural Fe_2O_3-Al_2O_3 ore or activated carbon as a catalyst.

351 Yushkevich, N. F. et al. J. Chem. Ind. (Moscow) 1933 (8), 50-8.
At 500-800°C hydrogen reacts almost completely with SO_2, but only 40-60% yield of S result H_2S being the other product. This H_2S can be converted to sulfur by reacting with SO_2 at 200-250°C using an iron ore or bauxite catalyst.

385 Pasheyskii, G. D. and Tabunov, K. A. J. Chem. Ind. (Moscow) 12, 480-3 (1935). Ural bauxite is a satisfactory catalyst for the reduction of sulfur dioxide to sulfur.

434 Farbaky, J. Bergh. Ztg. 1894, 234+.
The H_2S obtained by solution of the lead lining of the smelter in sulfuric acid is reacted with roast gases from pyrite by the Claus reaction. Granular sulfur is produced by spraying in a tower with calcium chloride solution.

436 Carulla, F. J. R. V. Soc. Chem. Ind., 16, 980-1 (1897).
A process for obtaining sulfur by the reaction of H_2S with SO_2 was discovered by accident when an old kiln was replaced by a new pyrites burner.

496 Jones, M.C. K. and Russell, M. E. Oil, Paint & Drug Reporter 123, No. 16, 30D (1933).
Under certain circumstances SO_2 and H_2S may be present in fuel gases and the Claus reaction might take place. This paper briefly considers this reaction, and outlines a method for the respective determination of H_2S and SO_2.

504 Applebey, M. P. J. Soc. Chem. Ind. (Chem. & Industry) 53, 1097-1101 (1934).
Deals mainly with the recovery of S from waste gases. In the Billingham Process, smelting gases are absorbed in an ammonium sulfite solution. SO_2, desorbed by heating, is reduced by passage over coke at 1100°C; by-products formed in the reducer are reacted with SO_2 over a catalyst.

571 Lee, J. A. Chem. Met. Eng. 49 (6), 80-82 (1942).
The Claus process is used to convert the purified H_2S from Arkansas sour gas to elemental sulfur.

PATENTS

34 Bolidens Gruvaktiebolag and Imperial Chemical Industries, Ltd. Fr. 847,325, (Oct. 6, 1939).
Catalysts for the Claus reaction are mixtures of portland cement or gypsum with pyrites.

42 Lindblad, A. R. and Herneryd, F. R. (to Imperial Chemical Industries, Ltd. and Bolidens Gruvaktiebolag). U. S. 2,200,928 (May 14, 1940).
A suitable catalyst comprises of pyrites or a pyrites concentrate and a water-absorbing material such as portland cement, smelt-cement or gypsum.

43 Lindblad, A. R. and Herneryd. F. R. (to Bolidens Gruvaktiebolag). Australian 109,531, (Jan 9, 1940).
The reaction is catalyzed by cement (portland or fused) or gypsum, to which are added metallic sulfides.

44 Imperial Chemical Industries, Ltd., Bolidens Gruvaktiebolag, A. R. Lindblad and F. O. Herneryd. Brit. 524,344 (Aug. 5, 1940).
Same as #34.

45 Imperial Chemical Industries, Ltd., Bolidens Gruvaktiebolag, S. J. Wallden and R. S. M Wallden (nee Sundgren). Brit. 524,345, (Aug. 5, 1940).
Claus reaction is carried out in the presence of catalysts which consist of water soluble alkali compounds, such as sodium carbonate, admixed with portland cement, aluminate cement, etc.

56 Wallden, S. J. and Sundgren-Wallden, R. S. M. (to Imperial Chemical Industries, Ltd.). U. S. 2,316,201, (April 13, 1943). Recovery of sulfur from gases such as metallurgical furnace gases containing H_2S, CS_2, and COS. The catalyst comprises a water soluble compound of an alkali metal, which forms an alkaline aqueous solution, such as Na_2O, Na_2S, or Na_2CO_3.

65 Fleming, E. P. and Fitt, T. C. (to American Smelting and Refining Co.). U. S. 2,388,259, (Nov. 6, 1945). The three step process for recovery of sulfur from roaster or smelter gases by use of a hydrocarbon fuel such as natural gas (1) Rapid reduction at 1200°C, (2) Conversion at 425-450°C as resultant product gases, (3) Exposing final gas stream to the Claus reaction.

70 Keeling, W. O. (to Koppers Co., Inc.). U.S. 2,413,714, (Jan. 7, 1947). The conversion to sulfur is improved by rapid cooling of the products of reaction with a water spray. Sulfur is removed as a liquid.

87 N. V. de Bataafsche Petroleum Maatschappij (W. F. Engel, inventor). Dutch 72,680, (July 15, 1953). Describes a catalyst prepared from Al_2O_3 and an aluminum salt, preferably $Al_2(SO_4)_2$ which gives higher conversions of H_2S and SO_2 to sulfur.

95 James, I. J., Jr. and McKinnis, A. C. (to Union Oil Co. of California). U.S. 2,756,127, (July 24, 1956). Oxidation reactions between gases ordinarily requiring high pressures can be carried out at low pressure by adsorbing first the less absorbed and then the greater on activated coconut charcoal with or without a catalyst. The production of sulfur from H_2S is cited as an example.

96 Conroy, E. H., Jr. et al (to Texas Gulf Sulphur Co.). U.S. 2,756,127 (July 24, 1956). Sulfur is produced from H_2S by oxidation and the Claus reaction. After sulfur is condensed from the gas stream, residual H_2S and S are oxidized to SO_2 which is absorbed in a liquid. This SO_2 is recycled.

99 American Smelting and Refining Co. Brit. 744,908, (Feb. 15, 1956). The tail gas from the Claus process is burned in a combustion zone and the SO_2 separated and recycled.

* 100 Shirimoto, Y. and Hoshiba, Y. (to Fuji Iron and Steel Co.). Japan 6615 (Aug. 7, 1956). H_2S for the Claus reaction is generated by reducing SO_2 by active hydrogen. The latter is formed by thermal decomposition of hydrogen gases.

103 Thumm, W. et al (to Metallgesellschaft A.-G.). U.S. 2,785,056, (March 12, 1957).
A series of four reactors, whose temperatures are changed periodically to drive off the sulfur from the coolest unit. Sulfur is removed from the gas stream by electrofilters during the changeover. The four temperature ranges for these units are 500-700°C, 350-450°C, 80-250°C, and 115-140°C. Activated alumina and iron oxide are the catalysts, carriers are aluminate cement, gypsum and kieserite.

104 Marsh, J. D. F. and Newling, W. B. S. (to Gas Council). Brit. 769,995, (March 13, 1957).
The Claus reaction is conducted continuously in a single reactor by recirculation of a mixture of waste gas and sulfur-free process gas to the reaction zone to hold the temperature down almost to the dew point of sulfur.

110 Bahr, J. U.S. 2,835,559, (May 20, 1958).
The unit apparatus for the production of sulfur from H_2S and SO_2 is described.

119 Piutsch-Bamag Akt.-Ges. (by A. Uffelmann). Ger. 1,051,359, (Feb. 26, 1959).
An apparatus for the Claus reaction is described. It consists of two parts, each part of which contains a catalyst chamber, a heat exchanger, and a washing arrangement.

125 American Smelting and Refining Co. (by E. P. Fleming and T. C. Fitt). Ger. 1,088,936 (Sept. 15, 1960).
The Claus reaction takes place at 400-600°C. The feed gas mixture is indirectly preheated by the product gases. After separation of the sulfur, residual H_2S in the waste gas is burned to form SO_2.

146 Power-Gas Corp. Ltd. (by D. A. Mitchell). Brit. 969,108, (May 2, 1952).
H_2S is removed from synthesis gas by adsorbtion in triethanolamine and iron or manganese oxides. H_2S is desorbed by heating the triethanolamine. SO_2 is formed when the iron or manganese sulfides in the second chemical absorption step are oxidized. SO_2 and H_2S are reacted via the Claus reaction.

230 Grekel, H. (to Pan American Petroleum Corp.). U.S. 3,317,281, (May 2, 1967). Deactivation of the bauxite catalyst for the Claus reaction is prevented by using an $H_2S:SO_2$ ratio of 2-2.5: 1, adding 0.5-1 mole% O_2 and preheating the gas mixture to 425-450°C.

267 Hall, W. A. U.S. 1,083,249, (Dec. 30, 1913).
Sulfur is produced from H_2S and SO_2 by bringing these gases together in theoretical proportions and maintaining them at 300-800°C until the reaction is completed.

284 Lewis, G. N. and Randall, M. J. Am. Chem. Sox. 40, 362-7, (1918).
Equilibrium for the reaction of H_2S and SO_2 was determined in a closed system at one atmosphere pressure (the boiling point of sulfur).

312 Werschen-Weissenfelser Braunkohlen A.-G. Ger. 514,570, (May 8, 1928).
Sulfur is prepared from gases containing H_2S by leading the gases over a contact agent of large surface in which excess SO_2 has been adsorbed.

324 Patentaktiebolaget Grondal-Ramen and N. E. Lenander. Brit. 352,477, April 5, 1930.
Gases evolved during the smelting of copper pyrite ores are reduced with coke at 350-450°C over a catalyst such as bauxite or Fe or Al oxide.

330 United Verde Copper Co. Ger. 562,517, (Sept. 23, 1930).
For this reaction, the catalyst contains a hydrated double silicate of Al or trivalent Fe and alkali or alkali earth metal, especially natural or synthetic zeolite.

341 Imperial Chemical Industries, Ltd. Fr. 717,934, (May 29, 1931).
The dust content of gases produced by the reduction of SO_2 with C is reduced to decrease the number of condensation nuclei. By slowly cooling the gas, large particles of sulfur are formed which are easily separated.

347 Baum, K. M. (to United Verde Copper Co.). Brit. 369,958, Mar. 29, 1932.
For the Claus reaction, the catalyst is a mixture of adsorbent materials, preferably activated carbon and a mineral gel, e.g., silica gel or a zeolite, in the ratio 2:1.

363 Baum, K. M. (to United Verde Copper Co.). Can. 332,208, (May 2, 1933).
Same as #347, the catalyst being activated carbon impregnated with a mineral gel.

364 Baum, K. M. (to United Verde Copper Co.). Can. 332,209, (May 2, 1933).
Same as #347, the catalyst being activated carbon impregnated with silica gel.

372 Ahlqvist, H. U. S. 1,955,722, (April 24, 1934).
The process is adopted for treating flue gases containing SO_2. The apparatus is described. The H_2S required for the reduction is made from the sulfur product.

373 Boswell, M. C. Can. 338,600 (Jan. 16, 1934).
The catalyst for the reduction may be oxides hydrated oxides, sulfides, hydrated sulfides or oxysulfides of a metal of the iron group which have been treated with H_2S at 200-300°C.

400 Boswell, M. C. U.S. 2,026,819 (Jan 7, 1936).
Same as #378.

407 Ahlqvist, H. Can. 356,673 (March 24, 1936).
Products from this reaction are cooled with a water spray, the volatile constituents are exhausted and the sulfur is separated and dried. Two thirds of the dried sulfur is treated with a hydrocarbon oil at high temperature to produce the H_2S required for the reaction.

435 Brock, J. and Hurter, F. Brit. 13,844, (June 23, 1896.
The Claus process wherein the H_2S is produced by treating alkali waste with lime kiln gases and the SO_2 is obtained by the combustion of iron pyrites or zinc blende.

468 Chemische Fabrik Rhenania and F. Projahn. Ger. 305,102, (1921).
The process using a bauxite catalyst at 150-300°C.

477 Kali-Chemie A.-G. (F. Projahn, inventor). Ger. 483,286, (June 8, 1927).
The residual gases from the manufacture of CS_2 are treated to recover sulfur by mixing them with SO_2 and passing the mixture at a raised temperature over contact masses of great surface such as bauxite, activated carbon or silica gel. The SO_2 may be prepared by burning a part of the gases with an excess of air.

485 Benner, R. C. and Thompson, A. P. (to General Chemical Co.). U.S. 1,773,293 (Aug. 19, 1930).
The process using dehydrated natural bauxite below 200°C.

486 Bacon, R. F. U.S. 1,782,226 (Nov. 18, 1930).
Iron pyrites is treated with steam at 600-700°C to form a gaseous product containing sulfur and H_2S. Oxidation of the residual iron sulfide in the pyrites produces SO_2. The H_2S and SO_2 thus formed react by the Claus reaction.

502 Baum, K. M. (to United Verde Copper Co.). U.S. 1,917,688, (July 11, 1933).
The Claus catalyst consists of particles of activated carbon impregnated with an artificial zeolite.

VII H_2S - LOW TEMPERATURE CLAUS:

60 Doumani, T. F. et al. Ind. Eng. Chem. 36, 329-32 (1944).
Duplication of previous abstract, (H_2 reduction, first card).

PATENTS

43 Lindblad, A. R. and Herneryd, F. O. (to Bolidens Gruvaktiebolag). Australian 109,531, (Jan. 9, 1940).
Sulfur is produced from H_2S and SO_2 at 120-300° being catalyzed by cement or gypsum to which are added metallic sulfides.

44 Imperial Chemical Industries, Ltd., Bolidens Gruvaktiebolag, A. R. Lindblad, and F. O. Herneryd. Brit. 524,344, (Aug. 5, 1940).
Same as #34.

54 Lindblad, A. R. and Herneryd, F. O. (to Bolidens Gruvaktiebolag). Ger. 721,053, (April 23, 1942).

85 Anglo-Iranian Oil Co. Ltd., J. N. Haresnape, and K. S. Cuddington. Brit. 689,664, (April 1, 1953).
Gases from petroleum-refining operations containing SO_2 (1 vol.), H_2S (2.1 vols.) and hydrocarbons (0-60 vols.) produce sulfur at 36-70% conversion by passing over bauxite or activated alumina at 120-160°C.

86 Anglo-Iranian Oil Co. Ltd., J. N. Haresnape, and K. S. Cuddington. Brit. 697,672, (Sept. 30, 1953).
A gas stream containing H_2S, SO_2 and hydrocarbons is passed over a bed of alumina or bauxite until the bed is saturated with sulfur, the catalyst is regenerated by removing the sulfur via distillation **or solvent extraction.**

97 Duecker, W. W. (to Texas Gulf Sulphur Co.). U. S. 2,767,662, (Oct. 16, 1956).
Sulfur is produced by the reaction of H_2S with SO_2 at temperatures below 150°C using alumina, bauxite, silica gel or activated charcoal as catalysts. Sulfur is removed from the spent catalyst by purging with a heated inert gas.

120 Metallgesellschaft Akt.-Ges. and Deutsche Gold-und Silber-Scheideanstalt vorm. Roessler (by W. Thumm et al). Ger. 1,064,925 (Sept. 10, 1959).
Alumina precipitated from sodium aluminate solutions by hydrolysis is a catalyst for the reaction of SO_2 and H_2S at temperatures below 150°C.

156 Urban, P. (to Universal Oil Products Co.). U.S. 3,149,920, (Sept. 22, 1964).
Sulfur produced by reacting H_2S and SO_2 over a catalyst containing 0.25-5% alumina in a silica base. Sulfur is removed from the loaded catalyst by dissolving the sulfur in a hydrocarbon such as xylene, benzene, or pentane.

196 Peter Spence & Sons Ltd. Belg. 661,331, (July 16, 1965). Gases containing up to 5% SO_2 are reacted with H_2S at low temperatures (about 100°C) over catalysts such as activated alumina without the reaction of oxygen. Sulfur is removed by heating with an inert gas at 445°C.

233 Peter Spence & Sons Ltd. Fr. 1,475,203, (March 31, 1967). Same as #196. The H_2S is produced by reacting a portion of the sulfur product with CO and H_2. Temperature range is 50-100°C.

239 Peter Spence & Sons Ltd. Brit. 1,097,306, (March 17, 1965). Same as #196. Temperature range is 100-150°C. H_2S is a by-product from a petroleum refinery.

297 Farbenfabriken vorm. F. Bayer & Co. Brit. 207,196, (Nov. 16, 1923).
Activated silica gel (Porous silicic acid) is the catalyst for the low temperature reduction. Sulfur is removed from the catalyst by extraction with Na_2S solution or an organic solvent such as CS_2, mono or dichlorobenzene or tetrahydronaphthalene. The sulfur is recovered by crystallization.

370 Rosenstein, L. (to United Verde Copper Co.). U.S. 1,941,623, (Jan. 2,).
The reaction is effected in the presence of an inorganic gel or adsorbent carbon and of water in the liquid state (suitably at a temperature above 120°C).

VIII LIQUID PHASE $H_2S + SO_2$:

55 Albertson, N. F. and McReynolds, J. P. J. Am. Chem. Soc. 65, 1690-1 (1943).
Numerous liquids are classified in which H_2S and SO_2 react quickly or slowly. The mechanism of the reaction is shown to be (a) The formation of compounds between the H_2S and the liquid medium through a H bridge (b) the release of H_2S ions by the addition compounds, and (c) reaction between SO_2 or its addition compounds and the hydrosulfide ion.

147 Averbukh, T. D. et al. Khim. Seraorgan. Soedin., Soderghashch, V Neft. in Nefteprod. Akad. Nauk SSSR, Bashkirsk, Filial 5, 115-22, (1963).
Experiments were performed with diethylene glylol, monoethylene glycol and triethylene glycol.

252 Klein, D. J. Phys. Chem. 15, 1-19 (1911).
The reaction of H_2S and SO_2 in a closed system is accelerated by the addition of small amounts of organic liquids such as ethyl alcohol, isobutyl alcohol, isoamyl alcohol, acetone, propyl acetate, benzaldehyde, carvone, methylethylketone, etc.

302 Matthews, E. J. Chem. Soc., 1926, 2270-3.
In the absence of a solid catalyst, it was confirmed that SO_2 and H_2S do not react with each other either in the gaseous or the liquid state except in the presence of a third substance which must be present in the liquid state.

331 Bary, P. Rev. Gen. Caoutchouc 8, No. 71, 25-6 (1931).
It is visualized that in the **Peachey** process the vulcanization of rubber with H_2S and SO_2 depends on the high dispersion of the sulfur since this is the dominant form of the sulfur produced experimentally by reacting SO_2 and H_2S in toluene.

333 Rao, B. S. and Rao, M. R. A. Nature 128, 413 (1931).
Solutions of H_2S and SO_2 in either benzene or carbon tetrachloride will react in the absence of water providing mercury is present.

440 Lang, W. R. and Carson, C. M. Proc. Chem. Soc. 1905, 158-160.
Laboratory experiments on the reaction of H_2S and SO_2 in the presence of moisture showed that sulfur and water were produced.

606 Makovetzkii, A. E. et al. J. Chem. Ind. (Moscow) 12, 362-6 (1935).
In the reaction of sulfur dioxide with water to form sulfuric acid and sulfur, the reaction rate increases with temperature and pressure. Since at high pressure the reaction should go at 150°, its commercial use is practical. Selenium is a catalyst for the reaction.

PATENTS

7 Rembashevskii, A. G. Russ. 42,545, (April 30, 1935).
Sulfur is obtained by the interaction of SO_2 and H_2S with concentrated sulfuric acid at 60-80°C.

10 Neesby, T. E. Dan, 52,359, (Nov. 30, 1936).
H_2S and SO_2 are reacted at 5°C or less in the presence of a solvent such as benzene, alcohols or water to produce colloidal sulfur.

18 Sulfur-Chemie A.-G. Brit. 459,418, (Jan. 7, 1937).
H_2S and SO_2 are removed from gases with the production of sulfur by introducing them into an agitated solution of thiosulfate. Thiosulfates may be prepared from these gases plus sulfites, sulfides, or polysulfides.

36 Gurevich, M. N. Russ. 55,771, (Sept. 30, 1959).
H_2S and SO_2 are passed through molten naphthalene and the sulfur formed is separated by treating the melt with organic solvents.

49 Wohlwill, M (to Metallgesellschaft Akt.-Ges.) Ger. 707,132, (May 15, 1941).
A flow sheet is given for reclaiming sulfur from H_2S and SO_2. They are passed through an alkylated aromatic base, e.g. dimethylaniline, which acts as a solvent for the sulfur. The excess SO_2 and the organic base are subsequently reclaimed for reuse.

68 Udy, M. J. Brit. 578,136 (June 17, 1946).
The process and equipment is described whereby H_2S and SO_2 are caused to react in an atmosphere of water vapor maintained over a body of aqueous liquid at a temperature of 50-100°C.

72 Udy, M. J. Can. 439,270, (Jan. 21, 1947).
H_2S and SO_2 are bubbled through or passed into a column of a solution of alkali metal sulfite to recover sulfur from these gases.

77 Ross, E. T. and Wilde, C. B. (to Stauffer Chemical Co.). U.S. 2,534,063, (Dec. 12, 1950).
Gaseous H_2S and SO_2 are mixed and bubbled into a water-acid solution wherein reaction occurs between the gases to form precipitated sulfur.

79 Wilde, C. B. (to Stauffer Chemical Co.), U. S. 2,562,158, (July 24, 1951).
H_2S and SO_2 at an $H_2S:SO_2$ ratio of 3-15:1 are reacted in water at 180°F to obtain a crystalline or microcrystalline form at sulfur rather than the usual amorphous or colloidal precipitate.

83 Udy, M. J. (to Guaranty Investment Corp. Ltd.). U.S. 2,619,407, (Nov. 25, 1952).
Same as #72.

111 Kali-Chemie Akt.-Ges. (by P. W. Schenk). Ger. 1,036,831, (Aug. 21, 1958).
Insoluble, plastic, fast-hardening sulfur is prepared by allowing SO_2 to react with H_2S in aqueous solutions (15% NaCl) in the presence of HCl (4%). The sulfur is precipitated by adding water.

126 Ignatyuk, V. G. USSR 138,233, (1961).
H_2S is oxidized with SO_2 in solvents, e.g. ethylene glycols. The sulfur suspension is separated, filtered, and air dried the clarified solution is regenerated by evaporation under vacuum.

132 Urban, P. and Glenn, W. K. T. (to Universal Oil Products Co.). U.S. 2,998,304, (Aug. 29, 1961).
SO_2 and H_2S react if they are absorbed in a H_2O saturated hydrocarbon containing an aralkyl sulfonate. For example, conduct the reaction in a cracked naptha containing SO_2, S, N, and H_2O at 112°F and 500 psia.

133 Badische anilin- & Soda-Fabrik A.-G. (by W. Fuchs). Ger. 1,091,995, Appl. (June 11, 1950).
H_2S is oxidized to sulfur with oxygen or SO_2 at 15-120°C at atmospheric pressure, by using an amide of the formula RCONR'R" as solvent, in which R, R', and R" are H, or alkyl, cycloalkyl or substituted phenyl or polymethylene residues.

143 Urban, P. and Massey, L. G. (to Universal Oil Products Co.). U.S. 3,023,088, (Feb. 27, 1962).
The reaction proceeds in a solvent consisting of a carboxamide and water. The carboxamide has the formula RCONR'R". R, R', and R" are either H or an alkyl halide.

144 Urban, P. and Massey, L. G. (to Universal Oil Products Co.). U.S. 3,050,370, (Aug. 21, 1962).
Sulfur is produced by absorbing SO_2 and H_2S in a solvent of methyl acetate, triethylene glycol, ethyl acetate, methyl proprionate, ethyl proprionate, or propyl proprionate containing 0.5 to 40 volume percent water.

145 Townsend, F. M. Brit. 895,503, (May 2, 1952).
H_2S and H_2O vapor are effectively removed from a gas mixture when it is contacted with SO_2 absorbed in an aqueous solution of an inert, neutral stable liquid organic solvent. The aqueous solution contains 90-99% of the organic solvent by weight. Characteristics of the organic solvent are described.

150 Urban, P. and Massey, L. G. (to Universal Oil Products Co.). U. S. 3,099, 536, (July 30, 1963.)
SO_2 and H_2S react in a liquid medium (H_2O-soluble aliphatic alcohols containing 10-40% H_2O) in which H_2S and SO_2 are soluble and sulfur is relatively insoluble. The reaction proceeds rapidly at ambient temperature causing the precipitation of crystalline sulfur.

197 Societe Nationale des Petroles d'Aquitaine (by P. Mathieu). Fr. 1,440,027, (May 27, 1966).
H_2S and SO_2 react in sea water at 50°C. The liquid is separated from the precipitated sulfur and neutralized to pH 6-7 before being recycled.

198 Institut Francais du Petrole, des Carburants et Lubrifiants (by P. Renault). Fr. 1,463,790, (Dec. 30, 1966).
H_2S from fuel gas is mixed with more than ten times its weight of $(CH_3)_2SO$ and the theoretical amount of SO_2, water content of the solvent is 1-40%. The water content can be controlled by evaporation. Typical conditions are 30°C and 15 atmospheres.

199 Institut Francais du Petrole, des Carburants et Lubrifiants. Neth. Appl. 6,604,053, (Oct. 3, 1966).
The reaction is conducted in an orthophosphoric acid ester, $PO(OR)_3$ where R=identicalent or different monovalent hydrocarbon (C=3 or more) groups, which has a solubility less than 1% at 20°C.

236 Deal, C. H., Jr. and Papadopoulos (to Shell Oil Co.). U.S. 3,363,989, (Jan. 16, 1968).
The selective solvent is sulfone, thus, a sour feed gas is passed into a reactor at 1000 psi and 80-250°F, where it is contacted with the absorbent reaction medium comprising sulfolane-95% H_2O-4%, $FeSO_4$-0.5%, and pyridinedicarboxylic acid (containing absorbed SO_2)-0.5%.

241 Feld, W. Ger. 202,349, (March 7, 1907).
Tar distillation products (so-called "heavy oils") constitute the reaction medium. The sulfur is soluble and the water can be conveniently vaporized. The formation of thiosulfate and polythianic acids is prevented.

244 Feld, W. Brit. 2,719, (Feb. 6, 1908).
Same as #241.

245 Field, - Fr. 387,438, (March. 6, 1908).
The solvent is not specified. The solvent evaporates off during the reaction and the sulfur crystallizes out on cooling.

251 Feld, W. Ger. 237,607, (April 18, 1909).
The reactants are allowed to act simultaneously or consecutively, upon moist suspended, or dissolved compounds (sulfite, thiosulfate, and polythionate, etc.) of metals whose sulfides (as those of Zn, Mn, and Fe) are insoluble and are decomposed by H_2SO_3 alone or in the presence of S, O, basic, neutral , or acid compounds, or such like substances.

256 Fritsche, P. Brit. 25,454, (Nov. 15, 1911).
Sulfur is produced by the action of H_2S on SO_2 in the presence of $Al(OH)_3$, or an Al compound of similar action, such as basic aluminum sulfate in solution, the gases being treated with liquid either simultaneously or successively. NH_3 may be present.

288 Harburger Chemische Werke Schön & Co. und W. Daitz. Ger 303,233, (Sept. 17, 1919).
The reaction chamber is a tower-like structure. SO_2 is introduced at the top, H_2S at the bottom and water is admitted as vapor, steam, or salt solution.

292 Ikeda, K. and Iwamura, A. Japan 36,316, (May 5, 1920).
The apparatus is composed of many small chambers, having rotating wings. Smoke containing SO_2 is passed into the chambers with a corresponding quantity of H_2S and exposed to continuous flow of H_2O by which SO_2 is changed to sulfur and water.

337 Bacon, R. F. U.S. 1,807,583, (June 2, 1931).
Reaction between H_2S and SO_2 is promoted by application of pressure (suitably about 5-10 atmospheres) to the reacting gases. Apparatus which employes violent agitation is also described.

361 von Szombathy, K. U.S. 1,936,959, (Nov. 28, 1933).
An arrangement of apparatus is described for conducting the crude gases into a polythionate solution. An excess of H_2S is used. The reaction temperature is 45-65°C.

368 Zorkoczy, G. Hung. 106,200, (May 1, 1933).
The gases containing H_2S or SO_2 are washed in an organic solvent miscible with water (e.g., Methanol, ethanol, glycerol, etc.) which contains the other gas in solution. About 2% of a sulfur solvent (benzene or toluene) is added to the mixture. The sulfur crystallizes in 98% pure crystals.

370 Rosenstein, L. (to United Verde Copper Co.). U.S. 1,941,623, (Jan. 2, 1934).
The reaction is effected in the presence of an inorganic gel or adsorbent carbon and of water in the liquid state (suitably at a temperature above 120°C).

402 Ward, A. L. et al (to United Gas Improvement Co.). U.S. 2,043,084, (June 2, 1936).
In reacting H_2S with SO_2 in the presence of an aqueous solution, a bath of the solution is swirled from below the surface upward and H_2S and SO_2 are simultaneously injected in jet form into the vortex of the swirl.

450 Feld, W. Ger. 272,474, (Dec. 31, 1911).
The two gases react with a solution of one or several metal thiosulfates. Any polythionate formed is reduced to thiosulfate. Thiosulfate strength is maintained by dilution.

474 Vogel, H. U. S. 1,546,048, (July 14, 1925).
In producing highly dispersed colloidal sulfur solutions, SO_2 and an excess of H_2S are simultaneously introduced into water in a reaction vessel.

498 Lincoln, B. H. (to Continental Oil Co.). U.S. 1,901,249, (March 14, 1933).
A portion of the H_2S is burned to form SO_2. The SO_2 is absorbed in water, and the aqueous solution thus formed is caused to react with the remainder of the H_2S to form sulfur.

500 Harrell, J. W. (to Standard Oil Development Co.). U.S. 1,915,364 (June 27, 1933).
Similar to #498 except that an excess of H_2S is used. After sulfur is separated from the residual gas, the residual gas is recycled to the combustion zone. Sulfur is removed from the aqueous slurry.

587 Every, R. L. and Grimsley, R. L. (to Continental Oil Co.). U.S. 3,318,666, (May 9, 1967).

587 (Continued)

A method of producing finely-divided dispersible sulfur which comprises passing a gaseous sulfur containing fluid comprising SO_2 and a second sulfur containing gaseous compound selected from the group consisting of hydrogen sulfide and mercaptans into an aqueous solution of 0.5 weight percent of an aliphatic alcohol of 1 to 5 carbon atoms.

604 Zorkoczy, G. Ger. 571,862, (March 6, 1933).
The solvent is methanol, ethanol, or glycerol containing 2% C_6H_6 or benzine, 1% toluene and 0.5% pyridine; sulfur is obtained in crystalline form.

628 Keller, J. L. (to Union Oil Company of California). U.S. 2,729,-543, (Jan. 3, 1956).
An aqueous absorbent is used. It is buffered to a pH between 4 and 6. The concentration of the thiosulfate ion is increased above the steady state value by adding a compound selected from the class consisting of water-soluble thiosulfates and water-soluble alkalis.

IX ALKALIZED ALUMINA:

121 Bienstock, D. and Field, F. J. J. Air Pollution Control Assoc. 10, (2), 121-5 (1960). A report of the bench-scale investigation of the removal of sulfur dioxide from simulated flue gas using alkalized alumina (metallic oxide absorbents). Data includes preparation of oxides and efficiency of absorption reaction.

148 Bienstock, D. et al. Am. Soc. Mech. Eng. - Paper 63-PWR-14 (1963). 6 p.
Cyclic process is described for removing oxides of sulfur from flue gas by adsorption at 625°F and then converting oxides to elemental sulfur by reduction of spent absorbent. Pilot-plant experiments are reported.

151 Bienstock, D. et al. J. Eng. Power 86 (3), 353-60 (1964).
Same as #148.

158 Bienstock, D. et al. J. Air Pollution Control 15 (10), 459-64 (1965).
Recent developments in SO_2 removal from combustion gases by dry techniques are reviewed. Process flowsheets, operating conditions, by-product yields and relative costs of a purification system for an 800-mw power plant are discussed.

201 Bienstock, D. et al. U.S. Bur. Mines, Rep. Invest. No. 7021, 52 pp. (1967).
A comprehensive report of the work described in #148.

635 Katell, S. and Plants, K. D. Hydrocarbon Process. Petrol. Refiner, 46 (7), 161-64 (1967).
A report compares capital requirements and operating costs for the alkalized alumina, Reinluft and catalytic oxidation processes. It also discusses Japanese processes and the uses of dolomite, limestone and phosphate rock as absorbents for SO_2 and SO_3.

638 Spencer, J. D. U.S. Bur. Mines, Inform. Circ. No. 8385 99 pp. 1968.
The removal of sulfur dioxide from gases is discussed. The alkalized alumina process is described and various other methods of removal are mentioned briefly.

X METALLIC SULFIDES:

253 Young, S. W. Mining Sci. Press, 103, 386-7 (1911).
The thiogen process consists of absorption of SO_2 by a metallic sulfide: $2\ CaS + 3SO_2 = 2\ CaSO_3 + 3S$ and the reduction of the $CaSO_3$: $2CaSO_3 + 2CH_2 = 2CaS + 2\ CO_2 + 2H_2O$

262 Wilson, F. L. Mining Sci. Press, 104, 497-8 (1912).
Experiments with the thiogen process are described. Flue gases rich in SO_2 (8%) are passed through an oil spray which in presence of CaS is changed to S and $CaSO_4$. CaS is made by roasting mixture of plaster of paris and sawdust in a combustion chamber.

281 Young, S. W. Met. Chem. Eng. 16, 309-10 (1917).
A description of the "dry thiogen process," experimental equipment and results is detailed. Assuming the use of petroleum as reductant, the approximate expression of the complex reactions is: $3SO_2 + 2CH_2 = 3S + 2CO_2 + 2H_2O$

289 Hutin, A. Rev. Prod. Chim. 23, 369 (1920).
A review of #281.

296 Wohler, L. et al. Z. anorg. allgem. Chem. 127, 273-94 (1923).
The formation of sulfur by the action of sulfur dioxide on calcium sulfide, zinc sulfide, and iron sulfide is discussed. The approximate reaction rates are summarized.

350 Schenck, R. and Hammerschmidt, F. Z. anorg. allgem. Chem. 210, 305-12 (1933).
The equilibrium constant for the reaction $CaS + 2SO_2 = CaSO_4 + S_2$ was determined. Isotherms are plotted at 850°C.

448 Same as #253.

451 Eddy, L. H. Eng. & Mining J. 93, 873-4 (1912).
Same as #281.

452 Same as #262.

453 Young, S. W. Eng. & Mining J. 95, 369-70 (1913).
Report of experiments with the thiogen process for trial runs made at the roasters at Campo Seco, California. The present status of the process is outlined.

457 Fulton, C. H. U.S. Bur. Mines Bull. No. 84, 94 pp. 1915.
This report covers general aspects of the smoke problem, gases and dusts in metallurgical smoke; recovery of flue dust and fume; removal of sulfur dioxide. The thiogen process is discussed.

458 Young, S. W. Am. Inst. Chem. Eng. Trans. 8, 81-89 (1915).
A summation of #281 and #453 for the thiogen process.

475 Milbauer, J. and Tucek, J. Chem. Ztg. 50, 323-325 (1926).
The chemistry of the roasting of some sulfide ores with SO_2 is discussed. The sulfides of the metals react at high temperatures with SO_2 according to the equation:
$RmSn + 2n\ SO_2 = Rm\ (SO_4)n + nS_2$

PATENTS

57 Mason & Barry Ltd. and E. Rich. Brit. 554,160, (June 23, 1943).
The process involves scrubbing gases containing SO_2 and other sulfurous compounds by means of a solution of Ca polysulfide in the presence of a catalyst to produce Ca polythionate. The calcium polythionate is decomposed into sulfur and hydrated calcium sulfate.

255 Sanborn, H. et al. Brit. 18,202, (Aug. 11, 1911).
Treating fumes containing sulfur oxides with a sulfide so as to form sulfur, sulfate, and sulfite. The sulfur is separated and the residue is reduced by carbon to sulfide for reuse. (The thiogen process). Details of construction and operation are specified.

259 The Thiogen Co. Fr. 450,786., (Nov. 20, 1912).
Same as #255.

263 Sanborn, H. et al. U.S., 1,062,120, (May 20, 1913).
Same as #255.

268 Young, S. W. U.S. 1,094,767, (April 28, 1914).
Removing sulfur dioxide and recovering free sulfur from sulfurous fumes by passing them through a heated chamber containing $CaCO_3$ or CaO and C or hydrocarbon vapor at a temperature of about 700°C.

307 Smaill, A. K. Can. 280,493, (May 29, 1928).
Sulfur is recovered from gases by segregating the SO_2 content, followed by a decomposing reaction with carbonaceous material and sulfide.

373 Rosenstein, L. and Klein, L. (to United Verde Copper Co.). U.S. 1,967,265, (July 24, 1934).
SO_2 is reduced by solid calcium sulfide at 700°C or higher. The calcium sulfate formed is then reduced in situ by the action of a gaseous reducing agent such as natural gas at a temperature above 700°C. This cycle is repeated and the sulfur produced is condensed and collected.

449 Same as #255.

643 Okunev. A. I. et al. U. S. S. R. 198,298, (Sept. 8, 1967.)
Reduction of SO_2 to sulfur by reaction with sulfides of alkali or alkaline earth metals at 400-1300°C.

XI MISCELLANEOUS METHODS:

113 Baccaredda, M. et al. Ann. Chim. (Rome) 49, 538-47 (1959).
Reduction of SO_2 with excess CH_4 on a kaolin catalyst at about 800°C. Principal products are CS_2, H_2S, Sulfur, H_2 and CO.

* 115 Berkman, Y. P. and Kushnir, S. V. Zhur. Priklad. Khim. 32, 1211-18 (1959).
Non catalytic reaction of SO_2 and CH_4 in a quartz tube below 800°C Ratio SO_2/CH_4 varied from 0.11 to 3.00. CS_2 is the product of interest.

* 117 Vdovichenko, V. T. et al. Zh. Prikl. Khim. 32, 347-50 (1959).
Studies of various catalysts on the reaction of SO_2 and CH_4 to form CS_2.

* 129 Golovei, M. I. Dokl. i Soobshch. Uzhgorodsk. Univ., Ser. Khim. 1961, No. 4, 33-4.
Role of quenching in the catalytic production of CS_2 from CH_4 and SO_2 suggests that catalyst initiates the reaction which continues in the gas after leaving the catalyst.

153 Pechkovskii, V. V. and Mal'tseva, T. G. Zh. Prikl. Khim. 37 (2), 240-6 (1964).
Reaction of CaO with SO_2 under reducing conditions. Carbon, H_2 and CH_4 were investigated as reducing agents. CaS, $CaSO_3$, and $CaSO_4$ were formed.

529 Avdeeva, A. V. J. Chem. Ind. (U.S.S.R.) 14, 1077-62 (1937).
Study of catalyst for the reaction of SO_2 with CS_2 and COS to form CO_2 and sulfur. Best results at 400-425°C on alumina.

313 Hunyady, I. and Koller, K. Hung. 102,324, (Aug. 24, 1928). Gas containing H_2S is reacted with an aqueous suspension of Mn_2O_3 to give sulfur and MnS.

406 The Gas Light and Coke Co. and R. H. Griffith. Brit. 449,710, (July 2, 1936). Covers catalysts for the reaction of SO_2 and CS_2 to form sulfur compounds of Ti, Th, U, Ce, Si, W or V separately or in mixtures are specified.

488 Levy, S. I. and Leaver, C. Brit. 358,851, (Nov. 5, 1931). Production of sulfur by the reaction of CS_2 or COS or both with SO_2 or O_2 or both on an iron oxide containing catalyst. Temperature 800-1000°C.

489 I. G. Farbenind. A.-G. (R. Suchy and O. Göhring, Inventors). Ger. 556,160, (March 15, 1931). P_2O_5 and sulfur are prepared by burning vaporized phosphorous in gases containing SO_2. Catalyst e.g. Na_2SO_3, Na_3PO_4 or KPO_3 is used.

625 Patton, R. A. (to Texas Gulf Sulphur Co.). U.S. 2,657,119, (Oct. 27, 1953). Similar to 641 below.

641 Produits Chimiques Pechiney-Saint-Gobain. Fr. 1,525,302, (May 17, 1968). SO_2 is reduced to sulfur with a hydroquinone compound. The quinone is reduced to hydroquinone and used in further SO_2 reduction.

642 Norddeutsche Affinerie. Japan 7682/68, (March 23, 1968). Removal of SO_2 from waste gas by treating with aqueous SeO_2 containing stoichiometric selenium.

XII REVIEW ARTICLES:

14 Secchi, I. M. Chimica e industria (Italy) 19, 576-8 (1937). Recovery of sulfur in flue gases.

26 Thau, A. Z. Ver. deut. Ing., Beiheft Folge 1938, No. 3, 81-6. Extraction and utilization of fuel gas sulfur (H_2S). Recovery as H_2SO_4, $(NH_4)_2SO_4$ or as elementary sulfur is described.

31 Lorenzen, G. Chem. Fabrik 1939, 6-23. Review of methods for removing dust, tar, NH_3, C_2H_6, NO and S from gases.

32 Nannestad, F. Tids. Kjemi Bergvesen 19, 109-14 (1939). Use of burnt pyrites as iron ore. Recovery of SO_2 from smelter gas.

38 Brabant. J. usines gas 64, 33-43, 49-56, 69-75 (1940).
Desulfurization of gas and processes for the recovery of sulfur.

47 Brownlie, D. Steam Engr. 10, 89-92 (1941).
Recovery of SO_2 from combustion gases.

91 King, J. G. Coke and Gas 16, 437-48 (1954).
Sulfur removal and recovery from coal and oils. Purification and sulfur recovery from fuel gases.

106 Bienstock, D. et al. U.S. Bur. Mines Inform. Circ. No. 7836. 96 pp. (1958).
Compilation of published information on chemistry emphasizing reactions having application to problems of air pollution.

116 Bienstock, D. et al. Natl. Acad. Sci.-Natl. Research Council Publ. No. 652, 54-62 (1959).
SO_2 air pollution is related to the sulfur content of fuels. A review of methods for removing sulfur from fuels and SO_2 removal after combustion.

127 Bienstock, D. et al. U.S. Bur. Mines Rep. Invest. No. 5735, 29 pp. 1961.
Survey of the magnitude of the SO_2 pollution problem. Report of bench scale studies on methods for removal of SO_2 from flue gas.

131 Simonin, J. C. Chaleur & Ind. 42, 251-66 (1961).
A review of atmospheric dispersion of stack gas and methods of removing SO_2 and fly-ash from flue gases.

137 Kettner, H. Lutverunreinigung 1962, 36-9.
The removal of SO_2 from the flue gases. A review of industrial as well as experimental methods.

142 Shibler, B. K. and Hovey, M. W. U.S. Inform. Circ. 8076, 62 pp. (1962).
Processes for recovering sulfur from secondary source materials. Volcanic sulfur deposits, H_2S from petroleum, SO_2 from flue gas and smelter gas pyrite and pyrrhotite.

155 Frankenberg, T. T. Am Soc. Mech. Eng. - Paper 64-WA/APC-2 12 p.
Discussion of the problems in removing and recovering sulfur before or after combustion of fuels.

159 Craxford, S. R. Ber. Int. Vortragstag. PRO AQUA, Basel, 1965, 77-92.
Air pollution from grit dust and SO_2. The dispersion of SO_2 from high stacks and removal of SO_2 from flue gas is discussed.

162 Frankenberg, T. T. Proc. Am. Petrol. Inst. 45, Sec. 3, 365-70 (1965).
Processes for removal of SO_2 from flue gas. Emphasis is on dry processes which have promise.

165 Johswich, F. Brennstoff-Waerme-Kraft 17 (5), 238-45 (1965) (Ger.)
Broad review of processes for flue gas desulfurization. Recovery as strong SO_2, H_2SO_4, and sulfur is covered. Non recovery control methods such as powdered dolomite injection is included.

166 Juentgen, H. and Peters, W. Staub 25 (10), 425-9 (1965) (Ger.)
Chemical engineering aspects for SO_2 removal from waste gas. Covers catalytic oxidation, activated carbon catalytic, "Reinluft," SO_2 adsorption on carbon and reaction with metal oxides.

167 Kettner, H. Bull. World Health Organ. 32 (3), 421-9 (1965)
Removal of SO_2 from flue gases. Wet scrubbing with H_2O and ZnO. Use of activated metal oxides. Solomite injection into boiler.

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169 Kriz, M. Sklar Keram. 15 (12), 374-6 (1965).
SO_2 removal from flue gases. A review of methods used in U.S.A., Great Britain and Germany.

170 Perry, H. Proc. Am. Power Conf. 27, 107-13 (1965).
Covers alkalized alumina, Reinluft and Pennelec catalytic conversion to H_2SO_4.

171 Rosendahl, F. Gas-Wasserfach 106 (31), 857-9 (1965) (Ger.)
Reduction of SO_2 using alunite as a catalyst indicates alunite is better than bauxite in the hydrogen reduction of SO_2.

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A brief review of flue gas treatment.

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Desulfurization of large quantities of gas. A discussion of the costs of various types of German installations.

314 Bruckner, H. Metallborse 19, 1352-3, 1464-5 (1929).
Technical processes for removing sulfur from gases. A review with bibliography.

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332 Kronacher, H. K. Gas Age Record 68, 37-43, 55-6 (1931).
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527 Anonymous. Chem. Eng. Min. Rev. 30, 49-51 (1937).
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Recovery of sulfur from smelter gas. A review.

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Removal and recovery of sulfur from smelter gas. A review of several absorption and reduction processes.

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Sulfur recovery from stack gases. A review including practices at Orkla-Grube, Norway, and Bolidens, A. B.

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Sulfur from industrial gases. A review.

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Review of processes for recovery or removal of H_2S and SO_2 from effluent gases. Catalytic conversion, solvent absorption applied to smelter and stack gases.

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285 Randall, M. and Bichowsky, F. R. V. J. Am. Chem. Soc. 40, 368-75 (1918). Equilibrium studies on the reaction between sulfur and water at high temperatures. Reference is not significant since modern data is available.

ADDENDA

646 British Patent 1,124,975 Derwent British Patent Abstracts, Vol. 8, #37 (1968), issued to Outokumpu Oy, Finland.

SO_2 gases produced from the roasting of sulfur containing ores and concentrates are reduced with a liquid fuel containing heavy hydrocarbons in the presence of free oxygen at 1000-2000°C. This first process stage is followed by two catalytic steps; the first at 400-450°C, and the second at 250-280°C.

648 T. D. Averbukh, N. P. Bakina, A. A. Radivilov, L. V. Alpatova, and E. A. Kravchenko. Khim. Prom. (Moscow) 44, No. 10, 753-7 (1968).

Experimental work on the preparation of elemental sulfur by the reduction of sulfur dioxide by natural gas (methane) is described in detail.

649 N. G. Vilesov and G. A. Gorbatych. Chim. Prom. Ukrainy, 2, 16-19 (1967).

Elemental sulfur is recovered from technical SO_2 gases by reduction with a gas mixture containing 90% methane, 2% hydrogen, 2% carbon monoxide and 6% inert (nitrogen). The process takes place at 800-1000°C using an alumina catalyst and a space velocity per hour of 180. The ratio of CH_4/SO_2 is 0.6 to 0.8.

647 Eng. Mining J., 169 (6), 91-100 (June 1968).

The sources of SO_2 leading to air pollution are discussed. Flow sheets and some cost data are presented for four processes used for recovering sulfur from stack gases; namely, the alkalized alumina process, the catalytic oxidation process, the activated char process, and the dolomite injection process.